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Optimal Design and Economic Evaluation for Integrated NGL/LNG Processes under Lean Feed Conditions

Lean Feed 조건에서의 NGL/LNG 통합공정에

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Abstract

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In recent years, the demand for lean gas fields has increased due to the development of unconventional gas reservoirs in Australia and U.S. Therefore, the re-evaluation of natural gas liquid (NGL) recovery and NGL/LNG integrated processes under lean feed conditions are required. This dissertation performs process optimization and economic evaluation for the various representative NGL recovery, natural gas liquefaction and NGL/LNG integrated processes considering the liquefied natural gas (LNG) higher heating value (HHV) specification.

Four different NGL recovery process schemes were evaluated under various lean feed conditions. The ISS and IPSI (A company name who owns it) processes are the representative processes in the conventional NGL recovery feed conditions. On the other hand, heavy hydrocarbons (HHC) separator and scrub column schemes are the simplified processes which may have advantages for lean feeds. The results indicate that IPSI process requires lowest raw material cost due to high process efficiency. However, its high total capital cost offset its overall economic

performance. The HHC separator scheme shows the lowest total capital cost because of a simplified configuration, but requires the highest raw material cost among the processes due to the poor separation efficiency compared with the other processes. ISS scheme shows best economic performance when the feed GMP value reaches 2.5. The scrub column scheme shows the best overall economic performance among the process schemes in the wide range of lean feed conditions. The results demonstrate that scrub column scheme can be seen as a good candidate of the NGL recovery processes for economically when the feed is in the considerably lean conditions.

An offshore platform has limited deck area different from onshore liquefaction plants. So the selection criteria for a liquefaction process is different compared to the onshore liquefaction processes. In this study, six types of liquefaction processes that applicable for offshore units were selected and analyzed both the process efficiency and economic performance. The six types of processes are a dual N2 expander, two single mixed refrigerant (SMR) and three kinds of dual mixed refrigerant (DMR) processes. The N2 expander process uses nonflammable pure nitrogen as the refrigerant that has advantages of safety and relatively simple operation. However, the liquefaction efficiency is the lowest one compared with the other processes. The DMR process includes two mixed refrigerant cycles that owns the highest liquefaction efficiency and per train capacity than the N2 expander and SMR processes. However, it has more complex process configuration than the other processes that normally used in the large scale liquefaction plants. On the other hand, the process performance of a SMR process is between the N2 expander and DMR processes. The results present that dual N2 expander process has the lowest process efficiency and net profit among the compared processes.

The SMR process 2 shows the lowest capital expenditure and payout time. The DMR base case scheme indicates the highest profitability and lowest operating cost among the processes because of the highest liquefaction efficiency. The results show that both SMR process 2 and DMR base processes have advantages in terms of some aspects compared with the other processes.

The proposed two simple integrated NGL/LNG processes and a conventional LNG and NGL coproduction process were also investigated with consideration of LNG HHV specification under the lean feed condition. The SMR process 2 and DMR base cycles were selected for the liquefaction part and the genetic algorithm (GA) method was used for the process optimization. The results show that the proposed integrated process, case 1, has overall economic advantages compared to the conventional base case scheme. The capital cost reduced remarkably by simplifying NGL recovery part, and only a little loss of liquefaction efficiency (less than 1%). The proposed process case 2, which adopts SMR process 2 as the liquefaction process, shows the lowest total capital cost and best profitability when a plant operating time is less than a certain period. Therefore, it could be a good process option when a plant reservoir lifetime is relatively short such as some peak shaving plants and special offshore applications in terms of economic aspect.

Keywords: economic evaluation, NGL recovery, liquefaction process, lean feed

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CHAPTER 1. Introduction

1.1. Research motivation

Natural gas is one of the representative fossil fuels that has a relatively low pollutant emission compared with traditional fossil such as coal and petroleum. Therefore, the demand for natural gas has increased in recent years. Natural gas is a globally used clean energy and important energy source which accounts for about a quarter of overall energy demand, especially the demanding of East Asia is increasing rapidly [1]. Recently lean natural gas production has increased since 2005 as growing production from shale gas reservoirs. With an increasing of unconventional gas (shale gas etc.) reservoirs in the area of Australia, East Africa and U.S, and demanding for cleaner energy resources in the world, the lean gas reservoirs exploitation are expected to continue to rise in the next few years [2], [3]. About 9.8% natural gas is supplied as LNG, and recently LNG demand growth is most outstanding in Asia countries [4]. Conventional natural gas reservoirs are usually normal or rich feed and most of previous studies were focused on conventional feed reservoirs. However, the unconventional sources like shale gas, tight gas occupy nearly 45% of remaining natural gas resources [5]. However, most previous studies regarding NGL recovery and LNG/NGL integrated processes were focused on the conventional feed (normal or rich) conditions. Therefore, re-evaluation of these processes under the lean feed conditions are necessary.

First, industry-standard single stage (ISS), gas sub-cooled (GSP) and IPSI are the representative NGL recovery processes. However, these processes are suitable for the conventional gas fields, which own much complex configurations in order to obtain high NGL recovery. Previous researches were mostly concentrate on improving recovery efficiency that the processes tended to become more complex. However, when a feed composition is lean, there is a possibility to simplify the NGL recovery process due to the little amount of heavy hydrocarbons. Additionally, a higher heating value (HHV) specification also was not considered by the most of previous studies. However, the minimum LNG HHV requirements for South Korea and Japan two countries are normally higher than U.S. and Europe countries. As a result, we design two simplified NGL recovery processes and compared the performance considering the economy impact as well as the LNG HHV specification with two conventional processes under various lean feed conditions.

Second, demand for natural gas liquefaction plants will also increase next few years as natural gas demand increases. Single cycle using a pure nitrogen or pure components cascade processes were widely used in the early stage. Liquefaction processes are complex and energy intensive that many advanced processes have been developed in order to improve LNG production and energy efficiency. Previous liquefaction studies normally concentrated on onshore plants and focused on improving process efficiency. However, an offshore liquefaction process has limited space that the selection criteria are different from onshore liquefaction process. There also some papers investigated liquefaction processes for offshore applications, but most of them did not consider capital expenditure or only analyzed a few specific liquefaction processes only. More complex configuration may obtain higher energy efficiency, but increasing equipment count also will increase total capital investment. Therefore, selecting a liquefaction process requires considering both capital cost and process efficiency especially in offshore units. This thesis selects various liquefaction processes, which are suitable for offshore application, and compared both efficiency and economy performance by each other.

Finally, both the NGL recovery and liquefaction processes require cold refrigeration systems that some NGL/LNG integrated studies have been conducted recently for sharing the refrigeration systems. Because integrating design could eliminate some equipment that have advantages over reducing total capital cost. Previous integrated studies focused on a normal or rich feed compositions and did not include the HHV specifications which are similar as the aforementioned NGL recovery researches. Therefore, in this thesis two simplified integrated NGL/LNG processes were proposed and compared the performance with a conventional integrated process considering HHV specification under the lean feed condition.

1.2. Research objectives

The purpose of this thesis is to analyze both process efficiency and economics of the representative conventional and proposed NGL recovery as well as NGL/LNG integrated processes, and to develop the economical processes considering LNG HHV specification under lean feed conditions. Additionally, this thesis quantitatively investigates the liquefaction efficiency and economy evaluation for the various liquefaction processes and suggests the suitable processes for offshore applications.

1.3. Outline of the thesis

Chapter 1 introduces the motivation and the main objectives of this thesis. In Chapter 2, investigates conventional NGL recovery and liquefaction schemes. The four NGL recovery and six types of liquefaction processes were simulated and compared the economic performance of each scheme under lean feed conditions. The total annualized cost was defined as an objective function considering capital cost, operating cost and profitability in NGL recovery. The processes were optimized the objective function by using a genetic algorithm. Both process performance and economic evaluation were performed on six types of liquefaction processes for offshore applications. The processes include one dual N2 expander, two types of single mixed refrigerant (SMR) and three types dual mixed refrigerant (DMR) liquefaction processes. Additionally, the previous studies regarding integrated NGL/LNG process was investigated. In Chapter 3 presents the proposed integrated processes. Two integrated NGL/LNG processes were observed in terms of process optimization and economic evaluation under the lean feed condition. Then, the proposed processes were compared the efficiency and economic performance with the conventional integrated process. Chapter 4 describes the main contribution of this thesis and suggests the future works.

CHAPTER 2. Economic evaluation of the conventional NGL recovery and liquefaction processes

This chapter is focused on the representative NGL recovery processes under lean feeds, and liquefaction processes for offshore units.

2.1. Evaluation of the representative NGL recovery processes considering LNG HHV specification*

2.1.1. Overview

Due to the higher value of heavier hydrocarbons as well as the necessary to maintain the pipeline specification for natural gas transportation, the natural gas liquid (NGL) are often extracted from original feed gas. From the previous research and industrial experiences, cryogenic processes are normally the most economical way to NGL separation. Industry-standard Single Stage (ISS) process is the representative process, which uses a turbo-expander for further cooling instead of Joule-Thomson (JT) valve [6]. Nevertheless, this process had some limitations such the relatively low NGL recovery and carbon dioxide freezing problem [7]. Thus, many advanced process schemes have been developed and introduced through papers or patents such as the gas sub-cooled (GSP), cold residue (CRR), IPSI processes etc. In addition, especially for lean natural gas feed condition the heavy hydrocarbon (HHC) separator and scrub column concepts were also introduced in some international conventions.

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* *The section 2.1 references the author's published journal paper: C. Jin, Y. Lim, Economic evaluation of NGL recovery process schemes for lean feed compositions, Chemical Engineering Research and Design 129 (2018) 297-305*

As the basic process, the ISS (Figure 2-1) has been normally cool the inlet feed gas using turbo expander and expansion valve. After passing a liquid vapor separator, the vapor portion is further expanded with turbo-expander and the liquid portion is expanded with expansion valve to operating pressure before introducing to the de-methanizer [6]. Because of its simple configuration and low CAPEX, many previous researches were carried out with the ISS process as one of the base case for further studies. [8], [9], [10]. The GSP scheme was first introduced by Campbell and Wilkinson. The GSP process is slightly different from ISS where the process dividing two gaseous streams after passing first separator. The first stream is expanded the same as ISS using turbo-expander before introducing to the de-methanizer and the other stream first heat exchanges with column top cold residue gas stream and after using JT valve for further expansion to column operating pressure. In the patent they proposed that the split vapor concept had an advantages in carbon dioxide icing in the column [6].

The IPSI process scheme (Figure 2-2) has the characteristics of high efficient and economical performance especially in separating propane, propylene and heavier hydrocarbon liquids which was developed by Yao(1999) . Unlike other NGL processes, the IPSI has focused on improving the tower stripping section. A portion of hydrocarbon liquid is withdrawn from one of the bottom tray and expanded and heated by inlet gas to produce a two phase, the vapor phase is recycled to the column increasing the light hydrocarbon component concentration and the efficiency of separation [11]. The previous research (Getu et al.) also maintained that the highest economic performance compared with other NGL recovery processes in some feeds composition [8].

The heavy hydrocarbon (HHC) separator and scrub column concepts are integrated NGL recovery process to the liquefaction unit and could consider for lean natural gas feed NGL recovery [12]. Typically scrub column and HHC separator could be used when natural gas feed is lean which has low C2+ contents, the processes have the advantage of lower capital cost than other NGL recovery schemes, but they have the limitation of low NGL recovery efficiency. Both scrub column and HHC separator remove heavy hydrocarbon after pre-cooling and prior to liquefaction process (refer to Figure 2-3 and Figure 2-4). Comparing with HHC separator, scrub column has less C1, C2 loss due to the effect of using a column instead of a separator. Even though they have introduced in U.S. patents, it is hard to find publications comparing HHC separator and scrub column economic performance with other NGL recovery processes [13], [14].

According to the previous NGL recovery studies, usually the conventional NGL recovery studies are most based on a normal or rich feed for comparing recovery performance or costs [15], [16], [17]. The ethane recovery plant in the South Pars gas field was simulated and analyzed with an advanced exergy analysis [18]. Retrofitting a NGL fractionation process was analyzed and proposed an energy efficient design in terms of side reboiler and heat pump hybrid system, they reported that the methodology could save considerable operating cost compared to the original case [19]. The existing NGL recovery plant in Sirri Island was analyzed with various feed condition and reported that with the increasing of heavier methane the work for compressors was reduced and cold heat exchanger showed the highest exergy destruction [20]. Maximizing a NGL recovery with the commercial simulation HYSYS and assessed the de-methanizer pressure [21]. An integrated NGL/LNG configuration was introduced and analyzed, the result showed a higher ethane recovery and considerable liquefaction efficiency [22].

Another integrated recovery plant was optimized several process parameters considering the effective operating conditions for plant performance with the object function of net profit [23]. A conceptual process design which integrate shale gas NGL recovery and LNG re-gasification in order to energy saving [24]. The dividing-wall column could replace de-ethanizer and de-propanizer columns in a NGL process and reported that this could enable weight reduction in floating liquefied natural gas facilities [25]. The hybrid genetic algorithms (GA) was used for optimizing economic problem of turbo-expander(ISS) recovery process [26]. Distillation system design was optimized for energy-efficiently separating multi-component mixtures for NGL processes [27]. Optimization of the CRR recovery process with the object target function of ethane recovery was performed and comparing with GSP and conventional turbo expander processes [28].

However, the increasing production of the lean feed requires re-evaluation of the economic performance of NGL recovery processes based on the lean feeds. In particular, the requirement of LNG HHV specification in East Asia is relatively high, and it weakens the advantages of high efficiency NGL recovery process. The specification of South Korea and Japan, market share of these two countries were more than 47% according to the IGU (international gas union) report [1], normally have a HHV specification between 39.7 and 43.5 MJ/Sm^3 which has a higher minimum value than U.S. and Europe [29]. Getu et al., 2013 showed economic performance of various NGL recovery processes for eight feeds composition but did not consider more simpler process such as HHC separator and scrub column concepts for lean feeds [8]. Ghorbani et al., 2012 used exergy pinch analysis for optimizing NGL recovery plants but limited to a normal feed [30]. Also the studies only focused on recovery efficiency or economic performance without considering the HHV specification. Park et al. 2015 includes both economic evaluation and HHV specification for comparing various patented NGL process schemes, but it remains only one normal feed condition [9].

In this chapter we focused on the four different NGL recovery processes with lean feed condition (GPM lower than 2.5) and evaluated its economic performance based on the total annualized cost (TAC). We also considered the LNG HHV specification which is required for the East Asia, such as South Korea and Japan.

2.1.2. Process selection and description

In this study we selected four types of NGL recovery schemes, the ISS, IPSI, HHC separator and scrub column, for economic evaluation. The ISS scheme is well known process in NGL recovery which leads to a major development in this field, after that emerging GSP, CRR, RSV and IPSI schemes and so on. Hence, selecting ISS as the base case here. Previous studies like Getu et al. 2013 and Park et al. 2015 analyzed ISS process [8], [9]. The IPSI is chosen because it was recently reported process for NGL recovery and Getu et al. 2013 reported that IPSI had good economic performance among various schemes [8]. The HHC separator and scrub column were not compared with the others even though they were introduced in the international conference [12]. Moreover, the proposed scrub column process in this thesis (Figure 2-4) is simplified version based on a US patent [31]. Because it is expected to have better economic performance in the lean feed composition, both of the processes are selected in this study. All the selected process models were developed by a commercial software Aspen HYSYS, and the Peng-Robinson equation of state was selected for the HYSYS simulation as its good prediction in hydrocarbon mixtures. Because it is expected to have better economic performance in the lean feed composition, both processes were selected in this study. All the selected process models were developed by a commercial software Aspen HYSYS, and the Peng-Robinson equation of state was selected for the HYSYS simulation as its good prediction in hydrocarbon mixtures [17].

- **Industry single-stage process (ISS)**

The ISS scheme process flow diagram is depicted in Figure 2-1. After removal of acid gas and water, treated feed gas stream at a pressure of 60bar and a temperature of 30°C pass the heat exchanger (E-100) and it is cooled down to about -33°C by gas stream 6, which is from the top product stream of the column (T-100). Then stream 1 is flashed in the separator (V-100) to vapor stream 2 and liquid stream 3. The vapor stream 2 is depressurized by the turbo-expander (K-100) to column top pressure and introducing to the column top stage, where the power energy generated by expander will be used to run the compressor (K-101). The liquid stream 3 from flash separator (V-100) is expanded by Joule-Thomson valve (VLV-100) to column pressure and introduced to the column (T-100). The column top product stream 6 is used to cool the feed gas stream and after passing the heat exchanger (E-100) the temperature rises to about 23°C. The residual gas stream 7 is compressed by the compressor (K-101 and K-102) to meet the specified natural gas pressure about 60bar, then cooled to specified temperature 30C by the heat exchanger (E-101).

The liquefaction section is not the main of this study because this study is focusing on the economic performance of NGL recovery section. Therefore, it was assumed that the total capital expenditure is a constant value in each scheme and does not affect the economic evaluation results of NGL recovery. The NG stream is entering a liquefaction process and temperature is reduced to about -160°C for producing liquefied natural gas. The stream 10 is further expanded to about 1bar by JT valve, the stream 11 after expansion may contain a small portion of vapor phase and separated through the end flash and the liquid stream is sent to the LNG storage.

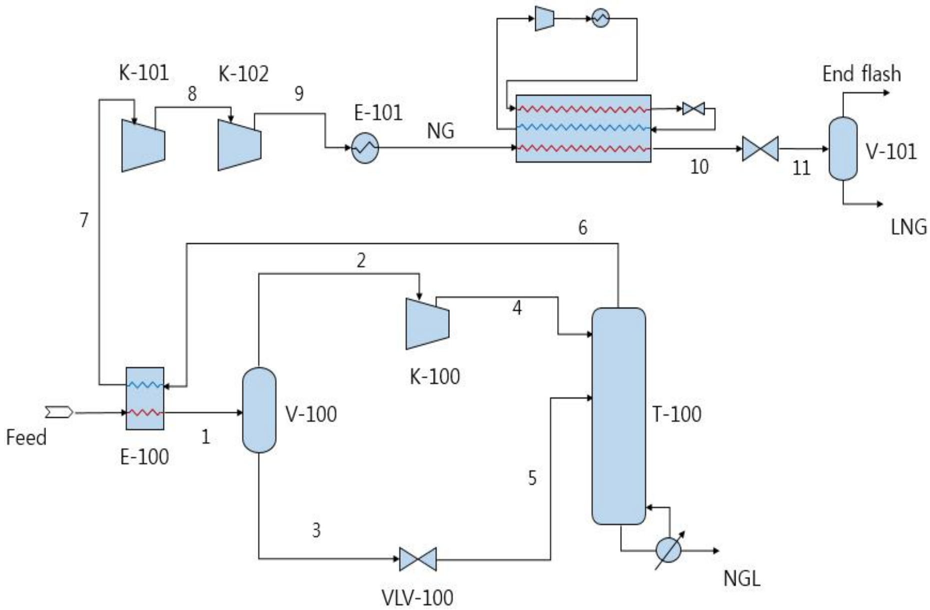

Figure 2-1 The ISS process scheme

- **Enhanced NGL recovery process (IPSI)**

The IPSI is recent patented scheme in NGL recovery field, and it uses the column side streams to improve ethane recovery without additional refrigeration system. Figure 2-2 shows the process flow diagram of IPSI process scheme. The cleaned feed gas stream is divided into stream 1 and stream 2. The stream 1 goes into the heat exchanger (E-100) and cooled by residue gas stream 16 then mixed with stream 6 before entering cold separator (V-100). The stream 2 is cooled via heat exchangers (E-101, E-102, E-103, and E-104) successively by the column side pump around streams PA1, PA2, PA3 after mixing into stream 8 at the temperature about -35°C. The stream 8 follows to the separator (V-100) and this stream is flashed into vapor stream and liquid stream 9. The liquid stream 9 is expanded via JT valve to the column pressure and introduced to the column whereas the flashed vapor stream is divided to two streams, where 30% is going into stream 11 and the other 70% to stream 12. The stream 11 is further cooled in the heat exchanger (E-105) by the column top product stream 15 and introduced as top feed stream 14 into the column T-100, the stream 12 is expanded via turbo expander (K-101), which can efficiently generate power for utilizing in compressor (K-102), to a temperature at -40°C before feeding into the column. The tower top product stream 15 is heated by the heat exchangers (E-105 and E-100) up to 24°C then compressed via compressor (K-102 and K-103) to the pressure around 60bar. The stream 19 is heated by compressing and should cool to 30C for meeting NG specification, via the cooler (E-106).

The major improvement of IPSI comparing with ISS or GSP is advanced utilization of the column bottom pump around streams. The pump around stream

PA1 and PA2 from T-100 bottom trays were used to cool the stream 5 and stream 4 then the warmed PA1’ and PA2’ returned to the column, respectively. These two pump around streams not only can reduce the column (T-100) re-boiler duty but also provide refrigeration to the feed stream as they get warmer after cooling stream 5 and stream 4. Stripping column pump around streams also can apply in ISS or other NGL recovery schemes. The considerable enhancement in IPSI scheme is made by using the PA3 stream. After stripping from the T-100 bottom tray the stream PA3 is divided into stream 20 and stream 21. The stream 20 is warmed in heat exchanger E-101 after cooling feed gas stream 2 and mixed with stream 27 then returned to the column bottom stage. On the other hand, stream 21 is reduce its pressure via expansion valve and heat exchange with stream 3 before entering the flash separator (V-101). The flashed liquid stream 24, which contains heavier hydrocarbon, will be pumped and mixed with column (T-100) bottom stream which later becomes the NGL product. However, the vapor stream 25 is compressed in K-100 to the column pressure and cooled by air cooler, then mixed with stream 28 before returning to the column bottom tray. It can decrease the required re-boiler duty because the PA3’ stream gets warmer. Additionally, relatively high concentration of light components in PA3’ stream the temperature profile inside the column also reduces and this makes increase heat integration ability [8].

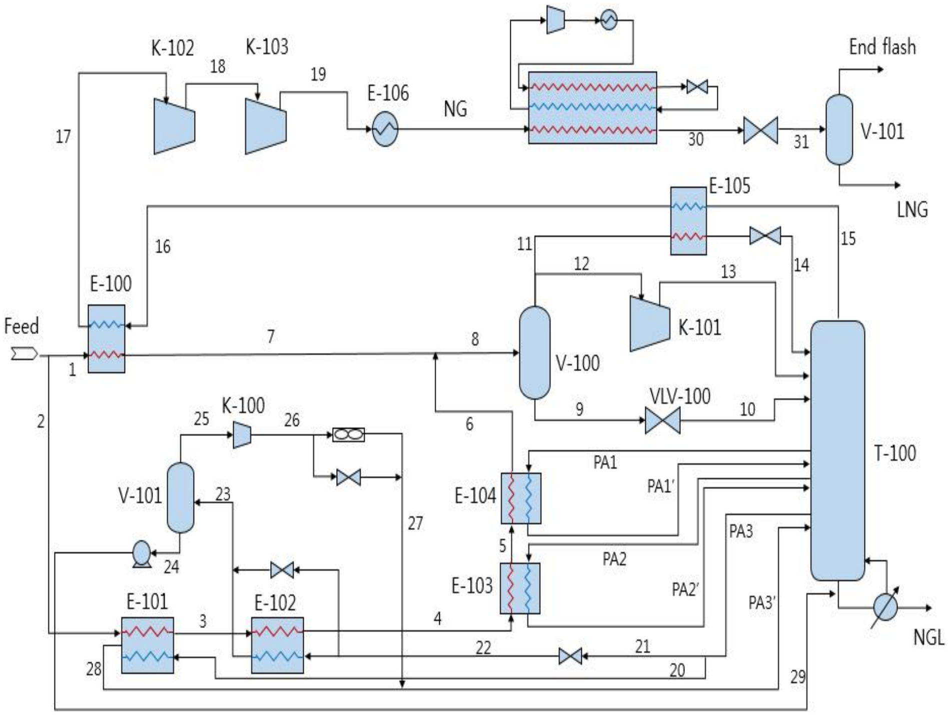

Figure 2-2. The IPSI process scheme

- **Heavy hydrocarbon (HHC) separator and scrub column process**

Unlike the mentioned NGL recovery processes which are independent with the liquefaction process, in the HHC separator scheme the treated feed gas is directly entering a liquefaction process and NGL recovery is performed during the precooling section in the main cryogenic heat exchanger. The HHC separator process flow diagram is shown in Figure 2-3. The stream 1 is stripping from cryogenic heat exchanger and adjust by E-100 exchanger for achieving optimum temperature then follows to the V-100 flash separator. The liquid stream is the recovered NGL stream, which contains relatively heavier hydrocarbons. On the other hand, the remaining vapor stream NG is reinjected to the liquefaction process for producing LNG product. The scrub column process is similar to HHC separator, but the only change is the column (T-100) instead of separator (V-100). Both HHC separator and scrub column schemes would have the advantages when the feed gas is lean because the complex recovery process such as ISS, GSP or IPSI would be inefficient due to small amount of heavy hydrocarbon contents while the HHC separator or scrub column concept (refer to Figure 2-3 and 2-4) can reduces capital cost.

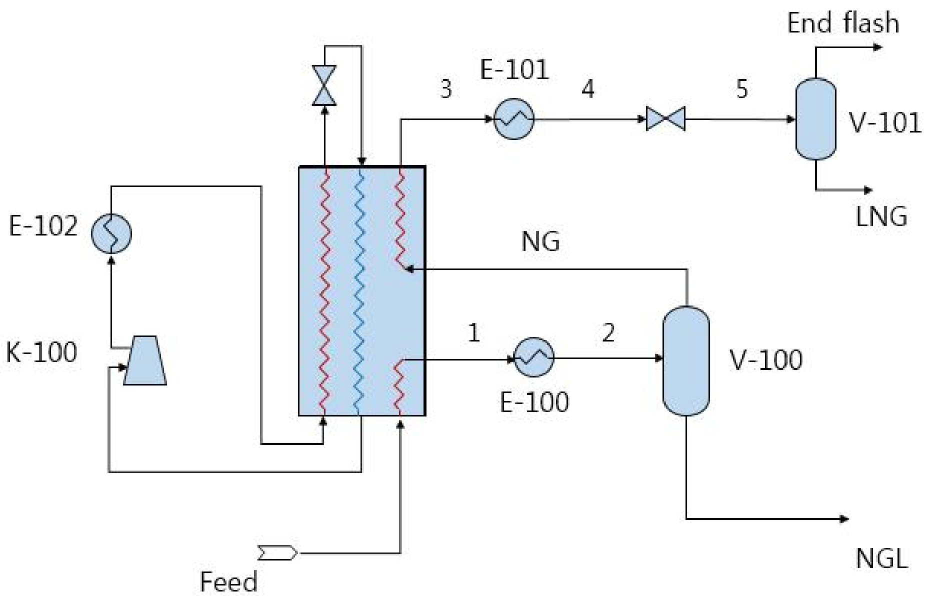

Figure 2-3. The HHC separator process scheme

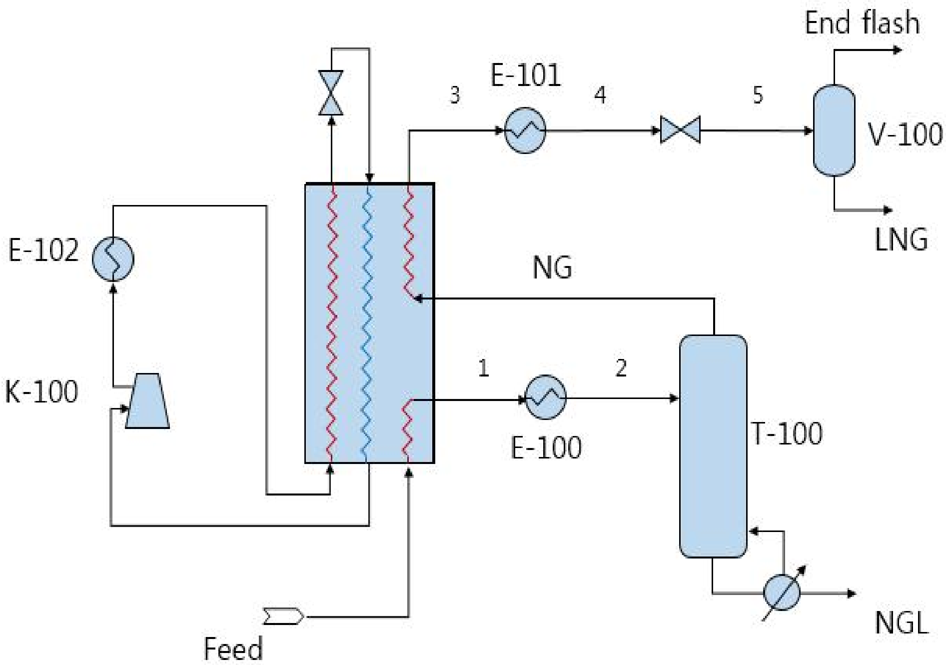

Figure 2-4. The proposed scrub column process scheme

2.1.3. Design criteria and specifications

A feed composition is an important and crucial factor on selecting a NGL recovery scheme [8]. In addition, the feed composition would vary as time passes. In order to cover the wide range of different lean feeds composition, four lean feeds having GPM lower than 2.5 are selected in this research as shown in Table 2-1. The GPM is defined as the amount of recoverable liquid expressed in gallons per 1000 standard cubic feet of a gas at 60F [32]. Additionally, the traditional NGL recovery process is after AGRU (Acid Gas Removal Unit) and dehydration process. Therefore, the feeds composition shown in table 2-1 is not include acid gas and water components.

To compare the different NGL schemes, specifying the key process parameters, which have a major effect on the plant performance, are very important. Thus, fixing these to the certain value and conditions are very significant for the faring comparison [8]. The specified key process parameters as shown in Table 2-2. The LNG production rate is fixed at 2MTPA (million tons per annum) for each scheme. The feed flow rate is adjusted to satisfy the constant production rate and the inlet feed pressure and temperature are kept as 60bar and 30°C. The compressor, turbine and pump efficiency is fixed at 80%. The pressure drop between heat exchangers was assumed as 0.5bar and 3°C of the minimum temperature approach was applied. The limitation of heavy hydrocarbon(C5+) concentration in LNG product should not exceed 0.1 mole % in LNG product due to freezing probability during the liquefaction process [33]. For meeting the specification of HHV, adding ethane or propane at liquefaction end is not preferable due to economy or technical reasons [34]. Therefore, it is better to adjust HHV at NGL recovery process. The HHV of LNG product is specified to 40.50 MJ/Sm^3. The same HHV value was used by Park et al. 2015 in previous studies [9].

Table 2-1. Feeds composition (mole %)

Component	Feed 1	Feed 2	Feed 3	Feed 4
C_1	92.97	91.57	91.22	90.67
C_2	4.18	5.18	5.21	5.23
C_3	1.23	1.5	1.68	1.98
iC_4	0.54	0.5	0.55	0.45
nC_4	0.54	0.48	0.52	0.65
iC_5	0.04	0.15	0.16	0.25
nC_5	0.04	0.1	0.11	0.23
C_6	0.03	0.01	0.01	0.02
C_7	0	0.01	0.01	0.03
C_8	0	0	0	0.02
C_9	0	0	0	0.01
C_{10+}	0	0	0	0.01
N_2	0.43	0.5	0.53	0.45
CO_2	0	0	0	0
GPM value	1.84	2.2	2.3	2.5

Table 2-2. Common process key parameters and constraints

LNG products	2 MTPA
Plant inlet gas pressure	60 bar
Plant inlet gas temperature	30 °C
Compressor/Turbine/Pump efficiency	80%
Pressure drop across the heat exchanger	0.5 bar
Heat exchangers minimum temperature approach	3 °C
HHV (LNG)	min. 40.50 MJ/Sm^3
C_4 (LNG)	2 mol% max.
C_{5+} (LNG)	0.1 mol% max.
Property fluid package	Peng Robinson

2.1.4. Economic evaluation

- **Capital cost estimation**

The major equipment of a distillation column, separator, plate fin type heat exchanger, shell and tube heat exchanger, compressor, turbine and pump are considered in the cost estimation. In order to estimate capital cost, sizing of each equipment is required and some parameters are obtained from HYSYS simulation results.

The column sizing requires the diameter and height calculation. The diameter of column calculation using Equation (2-1) introduced by Gavin Towler [35]. The equation is based on the well-known Souders and Brown equation which can get the maximum vapor velocity and the column diameter.

$$u_v = \left(-0.171{l_t}^2 + 0.27l_t - 0.047\right)\left(\frac{\rho_L-\rho_v}{\rho_v}\right)^{0.5},\ D_c = \sqrt{\frac{4V_W}{\pi\rho_v u_v}} \qquad (2\text{-}1)$$

where, u_v and l_t denote maximum allowable vapor velocity and plate spacing (range 0.5-1.5), respectively. V_w is the maximum vapor rate which can be obtained easily form the simulation. Distillation column height is calculated based on height Equivalent to theoretical plate (HETP) and the HETP=0.85m is selected here [35]. Separator diameter calculation use the gas flow rate Equation (2-2) which is applying droplet settling velocity [36].

$$q_s = 67824K_s d^2(1-M)\frac{P}{P_s}\frac{T_s}{T}\frac{1}{Z}\left(\frac{\rho_l-\rho_g}{\rho_g}\right)^{0.5} \qquad (2\text{-}2)$$

where q_s and K_s stand for gas flow rate and sizing parameter, respectively. The sizing parameter can be obtained either from American petroleum institute

(API) recommendation or from the droplet-settling equation, where we use a constant value 0.1 in this study. To calculate the heat exchange area of a heat exchanger, the overall heat transfer coefficient (U value) of 80 Btu/h ft2 °F is used [37]. The compressor, turbine and pump cost is relying on the power consumption of each equipment and the power duty is also can easily get from the simulation.

ESDU (Engineering Sciences Data Unit) organization provides the costing method of multi-stream plate fin type heat exchangers and ESDU 97006 was introduced about the selection and costing of plate-fin type heat exchanger [38]. The total exchanger volume should be obtained in order to cost the exchanger. First, a mean volumetric coefficient B_z is obtained from Equation (2-3).

$$\frac{\dot{Q}_z}{B_z} = \sum_{i=1}^{n} \frac{\dot{Q}_i}{\beta_i} \qquad (2\text{-}3)$$

where, $\dot{Q}_z$ is the heat transferred in the zone, n is the total number of streams involved in the zone, $\dot{Q}_i$ and β_i denote the amount of heat transferred to the i^{th} stream in the zone and the volumetric film coefficient at i^{th} stream. Next, the heat exchanger volume is then calculated from the expression below.

$$V_z = \frac{\dot{Q}_z/\Delta T_{m,z}}{B_z} \qquad (2\text{-}4)$$

where, $\Delta T_{m,z}$ is the logarithmic mean temperature in the zone. Finally, the total active volume of the exchanger is calculated by summing the volumes for each zone with considering an allowance of 15% for headers and distributers.

After sizing the equipment, purchase cost is estimated by using Equation (2-5), which is commonly applied for preliminary design stage [39].

$$log_{10}C_p^0 = K_1 + K_2 log_{10}(A) + K_3[log_{10}(A)]^2 \qquad (2\text{-}5)$$

where, C_p^0 is the equipment purchase cost. A is the capacity or size parameter for equipment. For compressors, turbines and pumps it represents power consumption. For heat exchangers and vessels, it represents area and volume, respectively. The correlation parameter data for K1, K2, K3 used in each equipment as below Table 2-3 which is taken from Turton et al., (2012)

The total capital cost (TCC) can be obtained by multiplying Lang factor with the major equipment purchase cost as well as considering time value which is based on the current CEPCI (chemical engineering plant and cost index) [40].

Table 2-3. Equipment cost data parameters

Equipment Type	Description	K1	K2	K3	Capacity, Units
Compressors	Centrifugal, reciprocating	2.2897	1.3604	-0.1027	power, KW
Turbines	Axial gas turbines	2.7051	1.4398	-0.1776	power, KW
Heat exchanger	U-tube	4.1884	-0.2503	0.1974	Area, m^2
Process Vessels	Vertical	3.4974	0.4485	0.1074	Volume, m^3
Pumps	Reciprocating	3.8696	0.3161	0.122	power, KW
Towers	Tray and packed	3.4974	0.4485	0.1074	Volume, m^3

- **Operating cost estimation**

Calculating the operating cost is often complex and it mainly rely on energy costs. In addition, utility costs like electricity are the main utilities and directly influenced by the cost of fuel price [8]. The utility cost, such as electricity and refrigeration cost, and the feed raw material cost are the main operating cost considered in this research. The raw material cost in here represents the cost of feed natural gas. The steam cost is ignored in this study because the column operating in similar conditions, and accordingly re-boiler temperature and column bottom NGL mass flow are also no big difference with all the candidate four schemes which result less effect in total operating cost. The electricity is mainly consumed by compressor, turbine and pump and the reference unit price of electricity and natural gas cost is obtained from the EIA (U.S Energy Information Administration).

Unlike ISS and IPSI, the HHC separator and Scrub column schemes should include additional refrigeration cost, extra precooling of NGL flow, for fair comparison of total operating cost because they get precooling refrigeration from integrated liquefaction system. The refrigeration cost is calculated by applying the liquefaction efficiency reported from previous references ([41]; [42]). The liquefactio efficiency in this study assumes 14 kW/ton/day and with the electricity cost of 0.1 usd/kWh, we could obtain the operating cost 33.6 usd/ton. Then multiply NGL mass flow rate for getting additional refrigeration cost for HHC separator and Scrub column schemes. In this study, we focused on NGL recovery part and neglect the heat exchanger cost effect of liquefaction cycle because the recovered NGL mass flow is very small compared to the LNG production (about 1.5% of LNG production) under the lean feed compositions. Thus, we ignore the increasing size of cryogenic heat exchanger in liquefaction part.

2.1.5. Profitability analysis

The profitability analysis is made for each scheme for comparing net profit and payout time. The net profit is obtained by the gross profit minus tax (the tax rate assumed 30% in this study). The gross profit is calculated by products revenue minus operating cost [40]. The payout time, also called payback time in some research, is evaluated and compared to each process. Payout time is the total capital cost divided to the net profit which is expressed as Equation (2-6) [43].

$$Payout\ time = \frac{Total\ capital\ cost}{Net\ profit} \quad (2\text{-}6)$$

2.1.6. Total annualized cost (TAC) optimization

The annualized TCC is expressed on annual basis by assuming 5% interest rate over 10 years' period for the economic evaluation. In this research, we considered TAC as object function of optimization. Here, the TAC is defined as the sum of annualized TCC, total operating cost and minus byproducts credits (Equation (2-7))

$$TAC = Annualized\ TCC + Total\ operating\ cost - Byproducts\ credits \quad (2\text{-}7)$$

In which the byproducts credits include each component of the NGL (C_{2+}) cost.

- **Global optimization**

The genetic algorithm (GA) was used by linking Matlab with HYSYS optimization to find the minimum TAC. The GA method has been used in some previous studies, and its applications have been tested [44]. GA is an intelligent random search algorithms based on the mechanics of natural evolution, it utilizes historical information and finally could yield a global optimum by repeating a series of population selection and generation for complex optimization problems which have many local optimum solutions. The parameter of population uses in this study is 30 and the elite count of 2 for the GA. The process constraints are given as above mentioned Table 2-2 and optimization variables by each scheme are listed in below Table 2-4. Where T, P and F stands for temperature, pressure and flow rate, respectively. The subscript number represents stream number of each process.

Table 2-4. Optimization variables by each scheme

Process schemes	Optimization variables
ISS	T_1, P_4
IPSI	F_{11}, F_{21}, T_7, T_3, T_4, T_5, T_6, P_{14}
HHC Separator	P_2, T_2
Scrub column	P_2, T_2

- **Sensitivity analysis**

The variable sensitivity analysis was performed for each scheme in order to investigate the effect of objective function.

For the ISS scheme (Figure 2-1), stream 1 temperature influence for the objective function TAC is less than 0.1 % when increase 10 % and decrease 10 % value compare to the base temperature value as shown in Figure 2-5. On the other hand, stream 4 pressure influence for the TAC is more than 1.5 % according to Figure 2-6. They show that the column pressure has bigger effect than the separator flash temperature in terms of the ISS process. Flow ratio of stream 11 and stream 21, the temperature of stream 6 and stream 7 were selected as the variables for the IPSI process (Figure 2-2). The flow ratio of stream 11,21 and the temperature of stream 6 do not have big impact (less than 0.2%) for the objective function TAC when increase or decrease 10% from the base value (refer to Figures 2-7~ 2-9). However, the flash temperature of stream 7 has bigger sensitivity in terms of TAC. The TAC difference is about 0.45% when stream 7 temperature increases 10%, whereas it has no feasible solution when decrease the temperature 10%. It shows the sensitivity for stream 7 temperature is big in the IPSI scheme. For the similar concept process HHC separator (Figure 2-3) and scrub column (Figure 2-4), flash temperature of stream 2 has bigger impact than the flash pressure according to Figures 2-10~ 2-13.

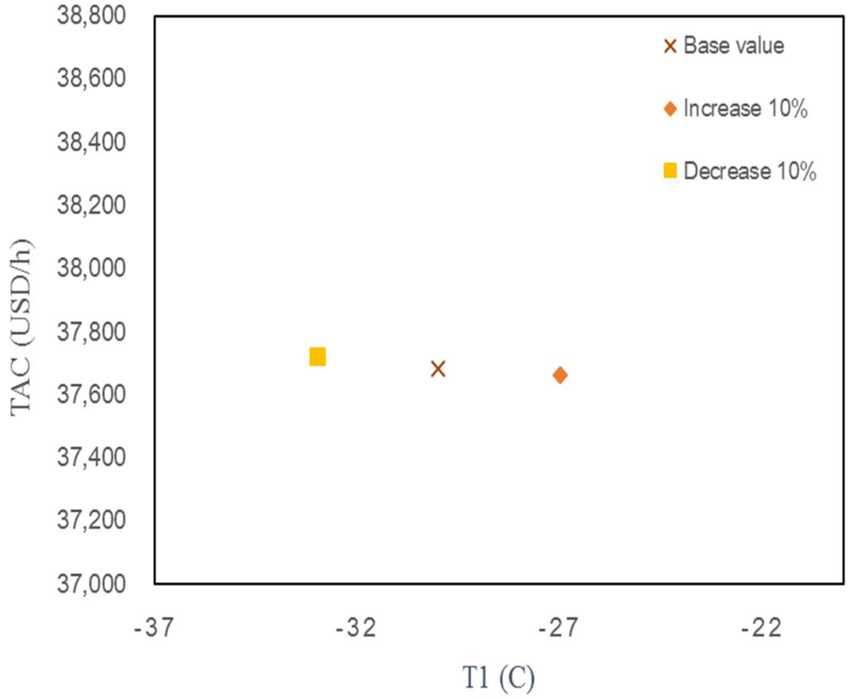

Figure 2-5. Temperature (stream 1) affect for the ISS scheme

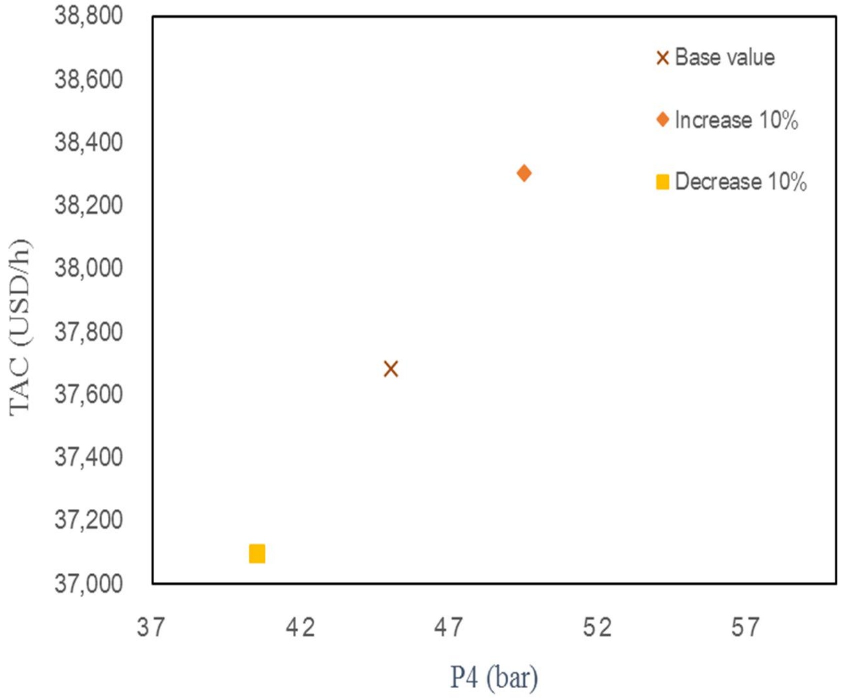

Figure 2-6. Pressure (stream 4) affect for the ISS scheme

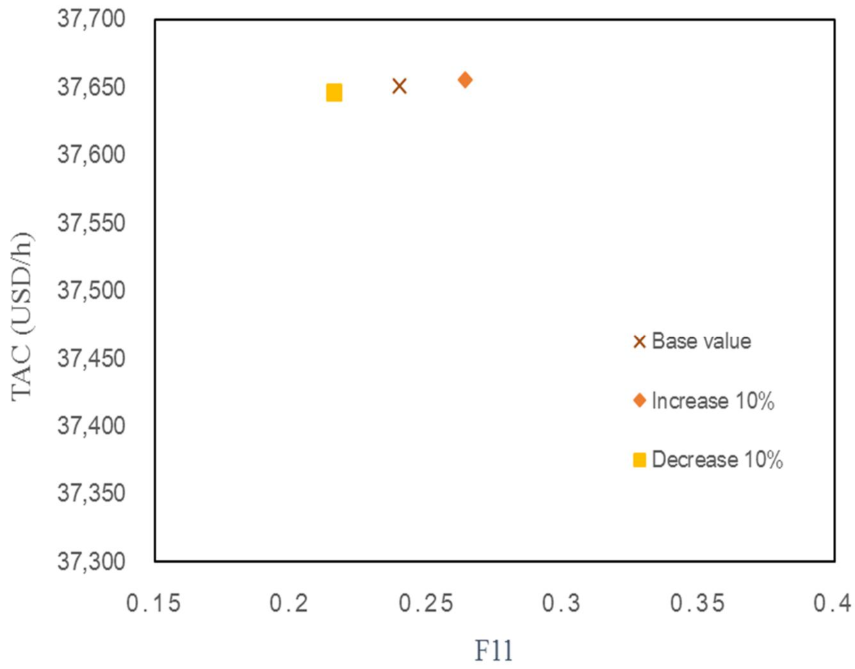

Figure 2-7. Flow ratio (stream 11) affect for the IPSI scheme

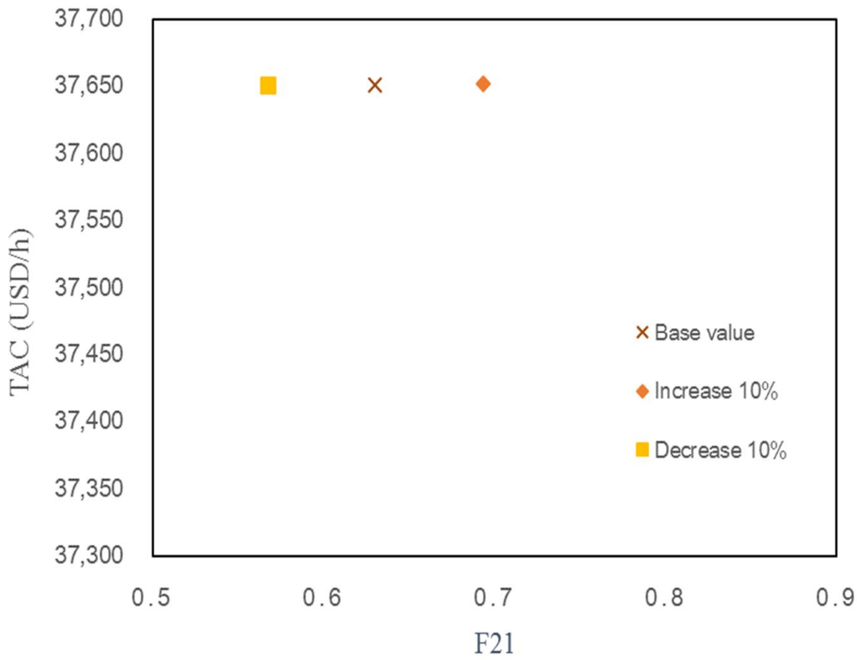

Figure 2-8. Flow ratio (stream 21) affect for the IPSI scheme

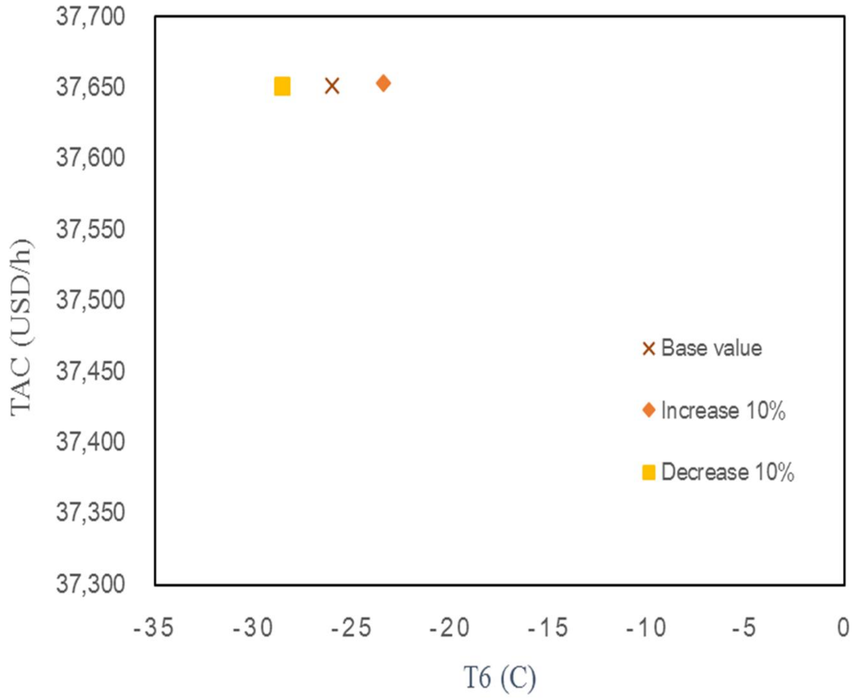

Figure 2-9. Temperature (stream 6) affect for the IPSI scheme

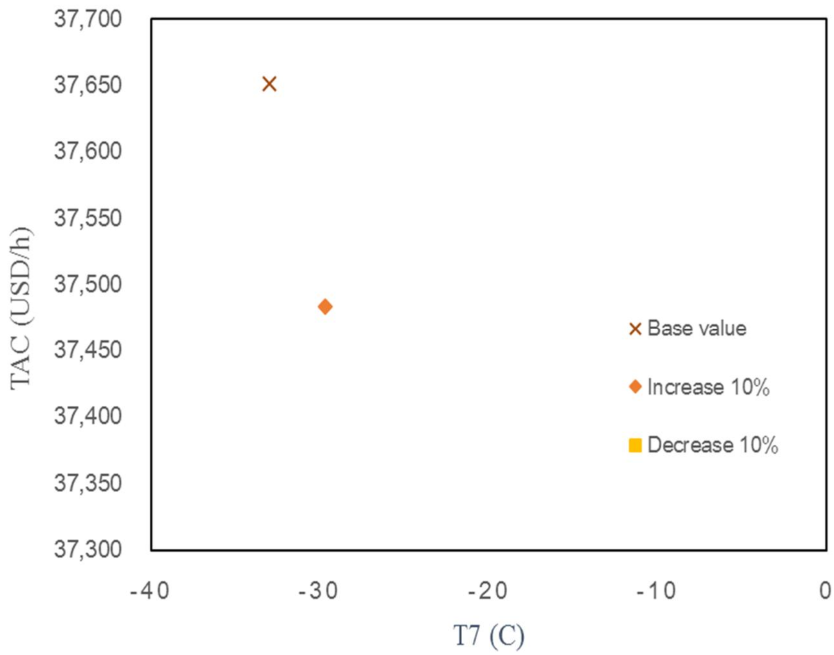

Figure 2-10. Temperature (stream 7) affect for the IPSI scheme

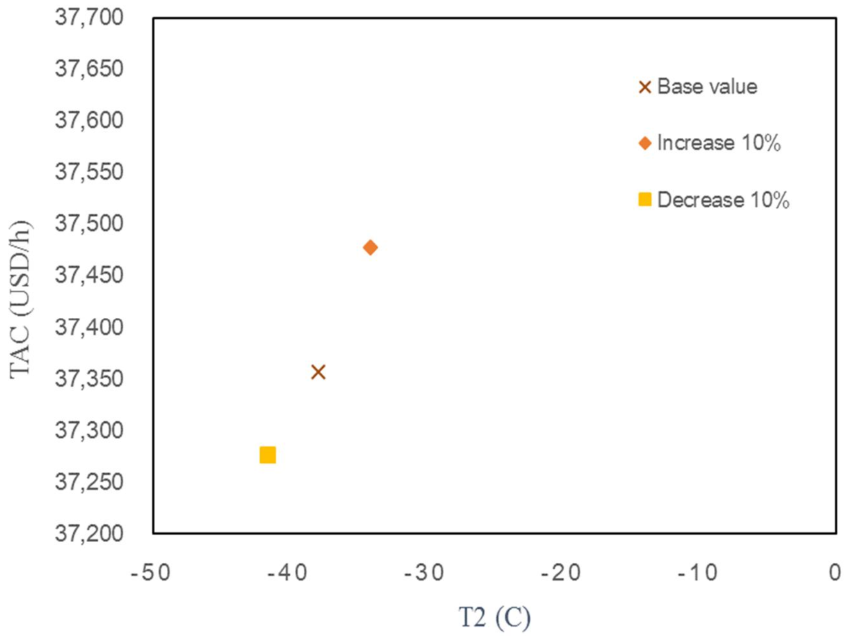

Figure 2-11. Temperature (stream 2) affect for the HHC separator scheme

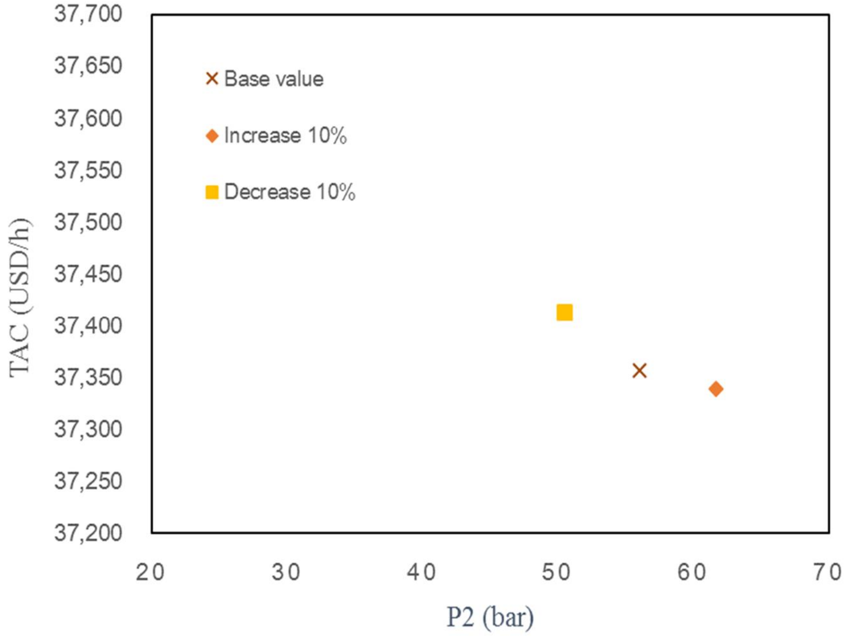

Figure 2-12. Pressure (stream 2) affect for the HHC separator scheme

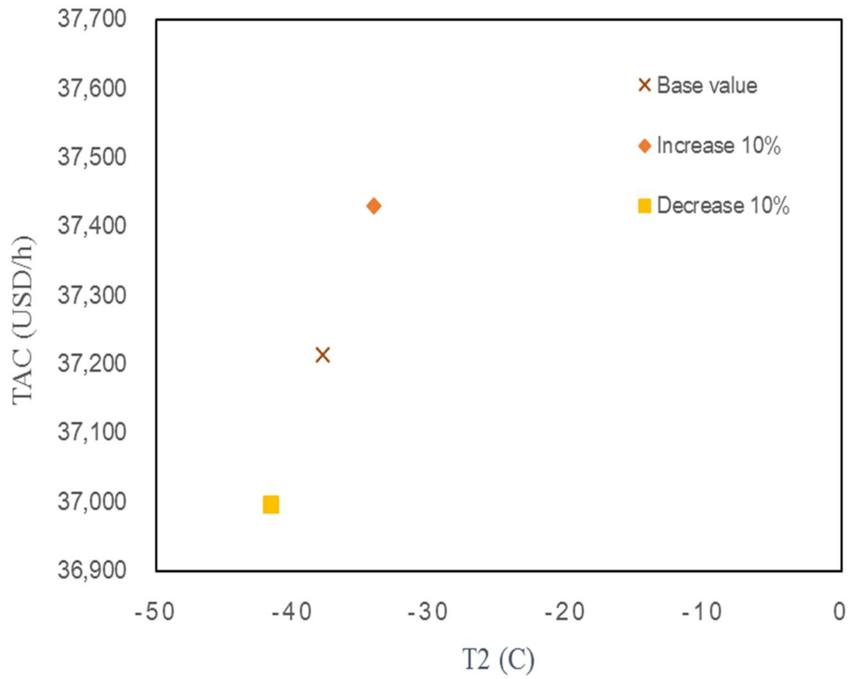

Figure 2-13. Temperature (stream 2) affect for the scrub column scheme

2.1.7. Results and discussion

The results of economic evaluation for the selected four schemes is shown as Table 2-5. The data in the table is taken the average values of the four different type feeds cases for comparison purposes.

The total capital cost result shows that IPSI has the highest value, whereas the HHC separator scheme has the lowest value among the other schemes due to its simple configuration and fewer number of process equipment. The most complex scheme IPSI gives 29.2% higher capital cost compared to the ISS scheme because the IPSI needs additional compressors and a number of heat exchangers for column bottom section improvements. Figure 2-14 illustrates the total capital cost by each scheme. The payout time presents similar tendency with the total capital cost results of the four schemes. However, remind that the payout time of each scheme is small according to Table 2-5. We also can find the payout time is similarly small according to the previous study conducted by Getu et al. 2013 [8]. This is because the payout time estimated by the academy usually assumes only the utility cost as the total operating cost due to the limitation of plant operation know-how. Therefore, the net profit overestimates the actual value that the results of payout is underestimated. If we consider other factors such as fixed charges and general expenses by a certain factor introduced in [40], the payout time of the ISS process is about 0.633, which is larger than the value shown in Table 2-5.

The total operating cost is mainly divided by the utility cost and raw material cost. The required raw material cost is illustrated in Figure 2-15. The IPSI process scheme requires the lowest raw material cost compared to other schemes even though it has the highest capital cost. The raw material saving for IPSI is 0.02% compared to ISS, 0.05% and 1.66% compared to the Scrub column and HHC separator scheme, respectively. Because this study focused on lean feed condition, the high efficiency of the IPSI is not so remarkable. The total capital cost affect is bigger than raw material cost. It means that even though the IPSI scheme requires lowest raw material, the higher capital cost and the operating cost offset the economic performance. HHC separator scheme owns the lowest total capital cost but the higher operating cost affect more in economic performance.

Table 2-5. Economic and profitability analysis for the selected NGL schemes

	ISS	IPSI	HHC Separator	Scrub Column
Total Capital Cost ($)	12,505,611	17,662,398	771,518	1,517,911
Total Operating Cost($/year)	344,977,737	345,229,183	352,179,175	345,885,252
1. Utility Cost($/year)	754,981	1,089,377	2,240,383	1,561,578
2. Raw material cost($/year)	344,222,756	344,139,807	349,938,792	344,323,674
By products credits ($/year)	29,705,007	29,672,069	31,070,406	29,508,476
Gross Profit($/year)	364,727,270	364,442,886	358,891,231	363,623,224
Net Profit($/year)	255,309,089	255,110,020	251,223,862	254,536,257
Payout time(year)	0.049	0.069	0.003	0.006

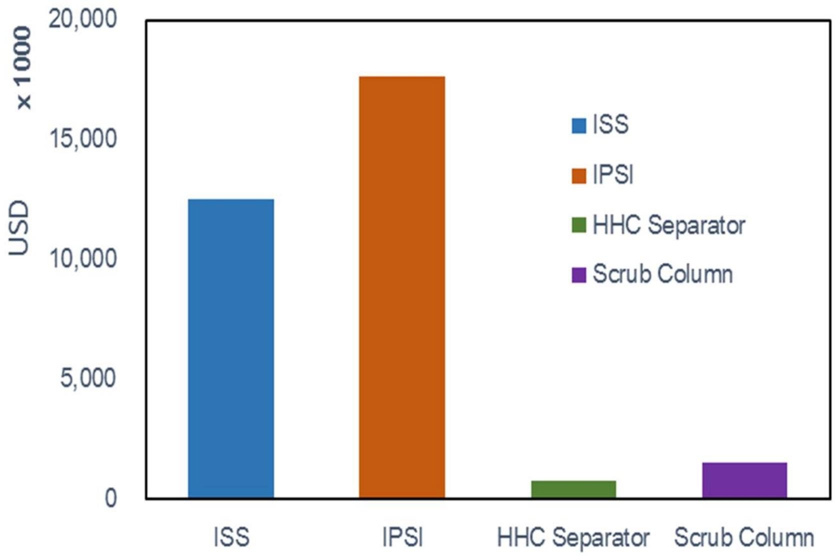

Figure 2-14. Total capital cost of each scheme

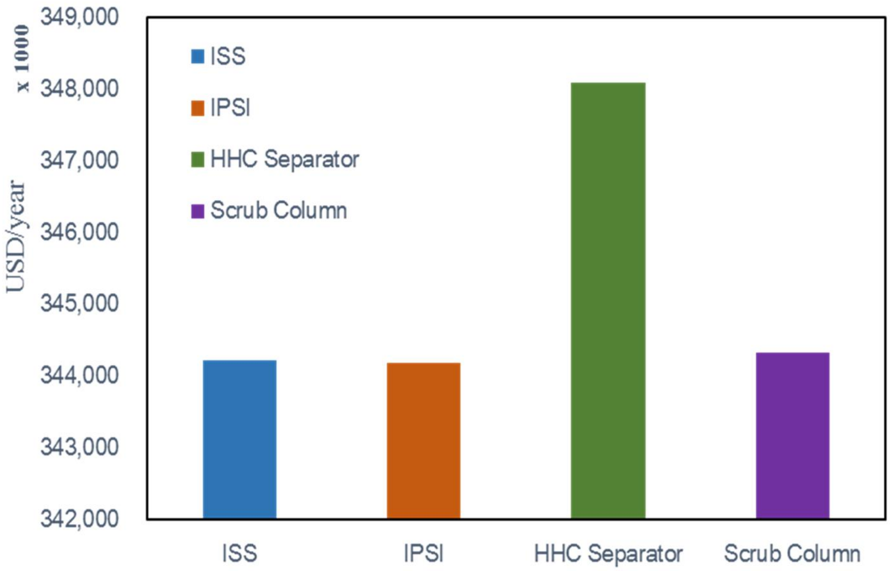

Figure 2-15. Required raw material cost by each scheme

According to the aforementioned economic analysis Table 2-5 the ISS scheme has better advantages of utility cost than others, however, the IPSI gives better results in raw material cost. On the other hand, the HHC separator has lowest total capital cost due to the simple design concept. Because of this dissimilarity in each scheme, we defined TAC for performing overall economic evaluation which was considering annualized total capital cost, total operating cost and byproducts credit as discussed in previous part in detail. The TAC value for each feed is shown in Figure 2-16. For the feed 1 (GPM 1.84) the scrub column scheme has the lowest TAC value and IPSI gives the highest TAC, the savings of scrub column scheme is 0.49%, 0.72% compared to the ISS scheme and the HHC Separator scheme, respectively. Similarly, for the feed 2 (GPM 2.2) the scrub column also has the best results and it is saving about 0.06%, 0.39% compared to the ISS and HHC separator schemes. Along with increasing feed GPM value the economy advantage of scrub column is lessened. When the GPM value reach 2.3 (Feed 3) both the scrub column and ISS scheme have almost the same TAC value, anyway they still have a lower value than IPSI and HHC separator schemes by 0.33% and 1.8% respectively. However, when the GPM value reach 2.5 (Feed 4) the ISS scheme, instead of Scrub column, has the lowest TAC value which is 0.32% lower than the scrub column and up to 2.62% lower than the HHC separator scheme.

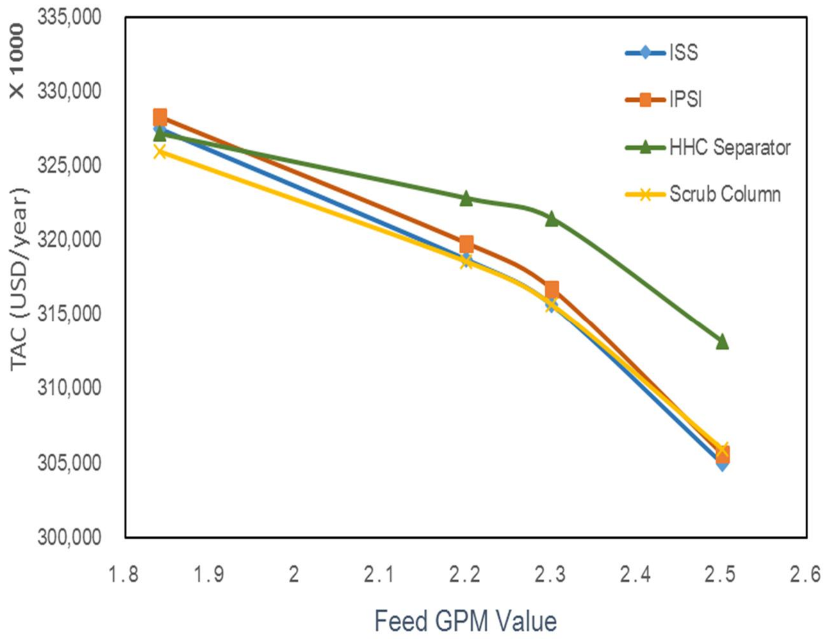

Figure 2-16. TAC analysis for each feed composition

2.1.8. Summary

The economic analysis was performed to the ISS, IPSI, HHC separator and scrub column NGL recovery schemes focused on the various lean feeds condition. The HHV specification was also specified for meeting the requirement of wide regions. The HHC separator scheme shows the lowest and the IPSI gives the highest total capital cost as usual due to the configuration complexity. The IPSI gives the best performance in required minimum raw material cost with respect to its high efficiency. Nevertheless, the high total capital cost and operating cost offset its economic performance. The HHC separator scheme has lowest total capital cost but shows the highest raw material cost due to its worst separation efficiency comparing to other schemes. Because of considering lean feeds, the required raw material cost shows not so much deviation among the schemes.

In this work, TAC was utilized as comparing overall economic performance for the selected NGL recovery schemes which was included the annualized total capital cost, operating cost and byproducts credits. The results show that the scrub column scheme has the best performance when the feed is lean and GPM is lower than 2.3. When the feed GPM value reaches 2.5 the traditional ISS scheme gives better performance. The results demonstrated that when the feed is considerable lean (GPM value is lower than 2.3) the scrub column scheme can be considered as a good candidate for a NGL recovery process.

2.2. Optimization and economic analysis of natural gas liquefaction processes for offshore units

2.2.1. Overview

LNG (Liquefied Natural Gas) has been widely used for economic long-distance transportation of natural gas because it has greatly reduced volume (about 600 times) compared to natural gas. Accordingly, demand for LNG plants are also expected to increase due to increased demand for LNG. In the early stage of liquefaction process, single cycle using pure refrigerant or cascade processes were widely used [45], [46], [47]. However, since the complex and energy intensive characteristics of liquefaction processes, many advanced liquefaction processes have been developed for improving production and liquefaction efficiency [48]. The representative processes are SMR, C3MR (propane pre-cooled mixed refrigerant), DMR and MFC (mixed fluid cascade). The C3MR process, developed by Air Products and Chemicals Inc., has remained largest market share because it has been dominant in land-based LNG plants. However, C3MR has no record for the application of offshore liquefaction units. It normally has large propane inventory which may increase safety concern and large amount of equipment count so that it is not suitable for offshore units.

Unlike an onshore liquefaction plant, an offshore platform has limited deck space, so that the selection criteria for liquefaction process is different with the land-based liquefaction plants. An offshore plant must be able to withstand much harsh environment than land-based plants such as waves or currents. It is very important to concern limited deck space, compactness and equipment count. Therefore, mixed refrigerant liquefaction process such as SMR, DMR and N2 expander processes are normally considered as the suitable processes for offshore

LNG process.

According to the previous liquefaction studies regarding offshore application, there are some focus on improving liquefaction performance. A SMR process was suggested for a small medium scale offshore unit because of small equipment count and simplicity [49]. A DMR liquefaction process was proposed that was suitable for LNG FPSO and showed that the proposed scheme decreased power consumption by 1.2% compared with a DMR process [50]. Four liquefaction processes such as a SMR and dual N2 expander processes for the small-scale plants were analyzed and showed that a SMR process had the lowest specific power requirement [51]. Optimization of a SMR liquefaction process for offshore applications was performed by knowledge-inspired hybrid approach [52]. The cascade, SMR and the single expander processes were simulated with Aspen HYSYS and optimized by the global optimization tool GA to achieve minimum power consumption [53]. N2-CO2 expander cycle, which has the inherent safety in the operation of offshore units, was optimized for compression energy requirements [54]. A cascade liquefaction process with nonflammable refrigerants was proposed for the offshore application and showed that enhanced efficiency compared to the conventional turbine-based processes [55]. A natural gas Claude cycle was analyzed and compared the efficiency and compactness with C3MR and some N2 Brayton cycles, showed that the Claude cycle was a good candidate for offshore LNG processes [56].

There also have some studies that both consider liquefaction efficiency and economic performance. Li and Ju, 2010. compared C3MR, mixed refrigerant cycle (MFC) and N2 expander processes and showed that N2 expander scheme was the best choice for offshore units in terms of efficiency and economically considering

LNG FPSO (floating production storage and off-loading units) layouts [57]. However, the efficiency and economic comparison of SMR and DMR cycles were not included.

Adopting more refrigeration cycles could increase the energy efficiency of a liquefaction process. But increasing the number of refrigeration stages also will increase equipment count and total capital investments. Therefore, selecting a liquefaction process requires a comprehensive comparison considering both capital cost and liquefaction efficiency especially in offshore units. In this chapter various liquefaction processes include N2 expander, SMR and some DMR cycles that suitable for offshore units were selected and comparing both the liquefaction efficiency and economic performance of each process.

2.2.2. Process description

A liquefaction process normally uses treated natural gas (NG) which is removed impurities such as acid gas, water, mercury and heavier hydrocarbons. The composition mainly contains C1(normally more than 90%) with small amount of C2-C4 components and very small amount of C5+ components (less than 0.1mole%). The treated NG is cooled to about -160C at standard atmospheric pressure through a liquefaction process.

- **N2 expander process**

A N2 expander process uses a pure N2 as refrigerant, it has advantages of simple configuration, relatively small footprint and easier operation than the other liquefaction processes. Thus, it can be applicable for some special offshore application and peak shaving plants. But it requires a lot of energy consumption that resulting in lower liquefaction efficiency than the processes using mixed refrigerant cycles. However, the N2 expander processes still favorable for the offshore units due to inherent safety in handling non-flammable refrigerant and easy operation. For example, the minimization of flammable inventory like propane is very important for process safety for floating liquefied natural gas (FLNG). Lee et al. 2013 analyzed several liquefaction processes using non-flammable refrigerants including single and dual N2 expander process [58]. A single N2 expander process has very simple configuration with one multi-stream heat exchanger, but the efficiency is almost lowest in the N2 expander processes. A dual N2 expander process has two expanders while the single N2 expander process has only one expander, so the dual N2 expander process has higher liquefaction efficiency than the single N2 expander process. The efficiency can be improved more than 32% [58]. Dubar 1988 introduced more complex processes: dual N2 expander and triple N2 cycles using several multi-stream heat exchangers [59]. Lim et al. 2014 simulated another configuration of single N2 expander process using two multi-stream heat exchangers [42]. The liquefaction efficiency is 679 kWh/ton, which was 14.5% higher and 26.2% lower efficiency than the single and dual N2 expander processes shown by Lee et al. 2013.

In this study a dual N2 expander process was selected and shown as Figure 2-17 due to its simple configuration and higher efficiency than single N2 expander process.

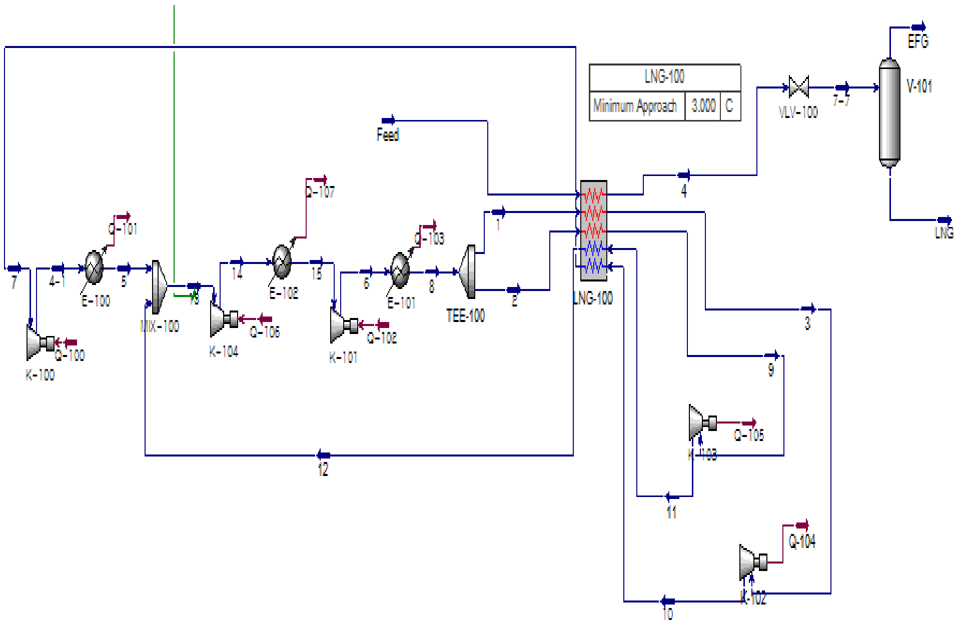

Figure 2-17 Dual N2 expander process

- **SMR processes**

For higher liquefaction efficiency, it is important to reduce the entropy generation because of temperature difference in the heat exchangers, normally mixed refrigerant (MR) cycle is effective in reducing the temperature difference with a lower equipment count than pure refrigerant cycle [60]. MR systems are widely used in liquefaction processes because of high energy efficiency, compact design compared to other processes operating with pure refrigerant. Therefore, a SMR process generally has higher liquefaction efficiency than a dual N2 expander process which uses pure N2 refrigerant only. The SMR process also can be seen as a good candidate that is suitable for the small scale offshore units because it can achieve higher efficiency than the dual N2 expander process and simpler configuration than DMR cycles. Moreover, SMR processes have higher flexibility and easier operability than DMR cycle because they use only one mixed refrigerant cycle. The capacity of a SMR process is normally less than 1.3 MTPA per train.

According to previous studies regarding the SMR processes, there have several different configurations. The representative SMR processes are PRICO SMR (Black & Veatch), APCI (air products and chemicals) SMR, Linde SMR. Khan et al. 2013 optimized a SMR by knowledge based optimization method [61]. The SMR process has very simple configuration which involves one multi-stream heat exchanger, one compressor, one set of cooler and one Joule-Thomson (JT) expansion valve. MR component they used N2, C1~C4 components and after optimization, the liquefaction efficiency was 424.3 kWh/ton (table 3-1). Shirazi & Mowla. 2010 also analyzed a SMR process and conducted energy optimization by using GA optimization technique [62]. Other researches from Morosuk et al. 2015 and Khan et al. 2012 also analyzed a SMR process [63], [64]. Their configurations for the SMR processes are almost the same except the numbers of MR compression stages (normally uses two or three compression stages). Xu et al. 2013; Cao et al., 2016; Lim et al., 2014 investigated a SMR process including a flash separator to separate vapor and liquid MR streams, after that the vapor and liquid MR streams increase the pressure by the compressor and pump, respectively [65], [66], [42]. The process could reduce compressor size because of using pump to increase a portion of liquid MR stream. Vatani et al. 2014 analyzed an APCI SMR that the process configuration consists of two multi-stream heat exchangers [67]. Moein et al. 2015 analyzed another APCI SMR process with three multi-stream heat exchangers, they optimized total required power of the SMR process by GA [68]. The MR compositions include N2, C1~C3, nC4, iC4 and specific power consumption of the process was 275.04 kWh/ton after optimization. There also exist more complex configurations than previous mentioned SMR processes such as Linde SMR process investigated by Vatani et al. 2014 [67]. The Linde process used four multi-stream heat exchangers in order to improve heat integration. However, this process has more complex configurations than APCI SMR and the efficiency advantage is not obvious, so does not include in this study.

As mentioned above, there are various versions of SMR liquefaction processes, from the simplest SMR process to the more complex Linde SMR process. This study we choose SMR process 1 (Figure 2-18) and SMR process 2 (Figure 2-19) as the candidate processes for the offshore applications considering simple operation and footprint.

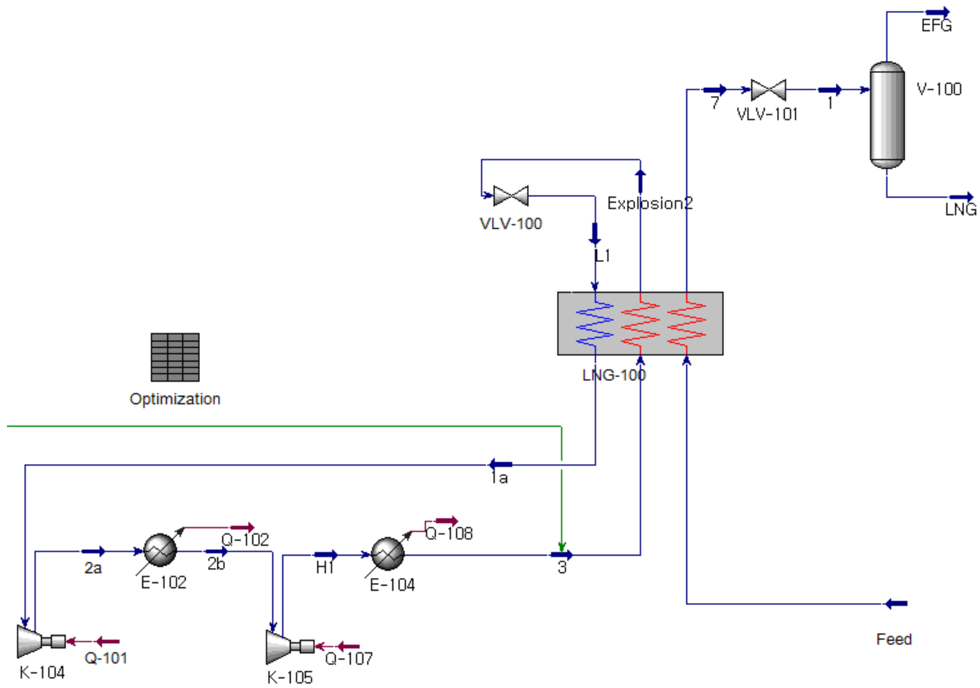

Figure 2-18. SMR process 1

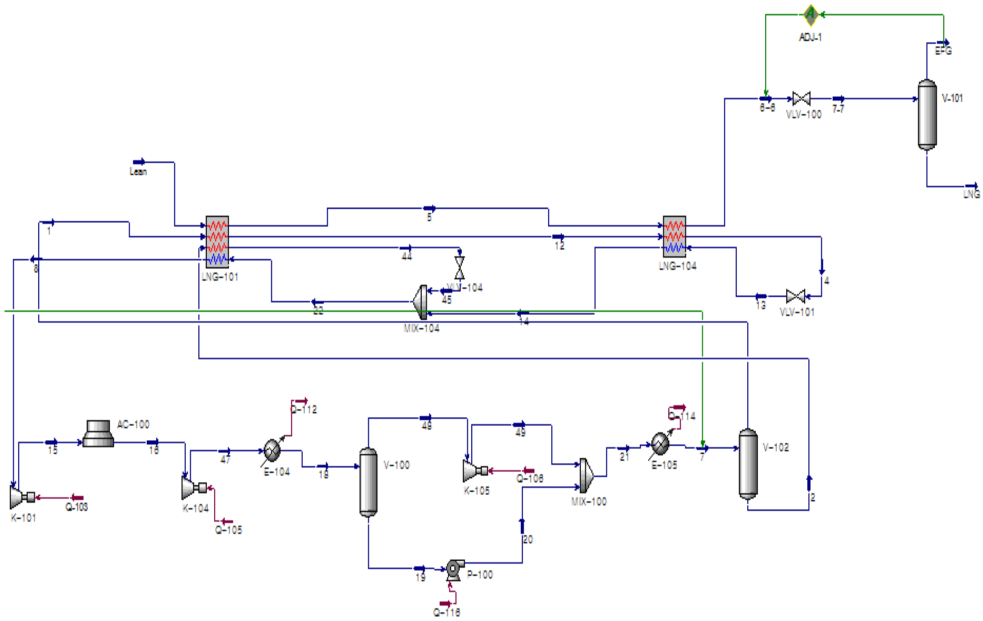

Figure 2-19. SMR process 2

- **DMR processes**

A SMR cycle has simple structure with less equipment count, but the efficiency and each train capacity are limited, so it has been used for the small size of NG liquefaction plants. Since a SMR cycle has limitations in liquefaction efficiency and each train capacity that can be improved, DMR cycle should be used for achieving higher liquefaction efficiency and larger capacity per train. The capacity can be achieved up to 5 MTPA per train for DMR cycle [69]. Therefore, the DMR liquefaction process could be applicable in large scale offshore units. For example, Shell Prelude FLNG adopted a DMR process. The previous studies Venkatarathnam & Timmerhaus, 2008; Vatani et al. 2014 investigated a conventional DMR base process that uses two multi-stream heat exchangers for pre-cooling and the other two multi-stream heat exchangers for liquefaction and sub-cooling successively [60],[67]. Another configuration of DMR process was simulated and economic optimization was performed by Wang et al. 2014 [70]. This DMR process uses only one multi-stream heat exchanger as the pre-cooling section, which is different with the abovementioned DMR process using two multi-stream heat exchangers for pre-cooling. Additionally, US patent 6,269,655 introduced several DMR liquefaction processes requiring minimum plot plan area that were suitable for offshore applications [71]. The invention addressed that these DMR processes could operate at high efficiency that were both compact and cost effective. Number of MR cycles is one of the most major determining factors for liquefaction efficiency, and increasing the number of MR cycles can improve process efficiency and capacity but the total capital cost and required area may increase accordingly. For example, the Linde introduced a liquefaction process called MFC using three MR cycles. The MFC process has higher energy efficiency than above introduced liquefaction processes due to uses three different MR cycles. However, the complex configuration and require large footprint that are not suitable for offshore application. As mentioned, compactness and simple operation are more important for offshore units. Though, the processes have much more complex configurations than DMR cycle such as MFC are excluded in this study.

The well-known DMR base process shown as Figure 2-20 and the other two cases liquefaction processes (Figure 2-21 and Figure 2-22), which were suggested for offshore applications by the US patent 6,269,655 [71], are selected as the candidate processes for DMR cycle. The liquefaction efficiencies for N2 expander, SMR, DMR processes referenced from the previous studies are listed in Table 2-6.

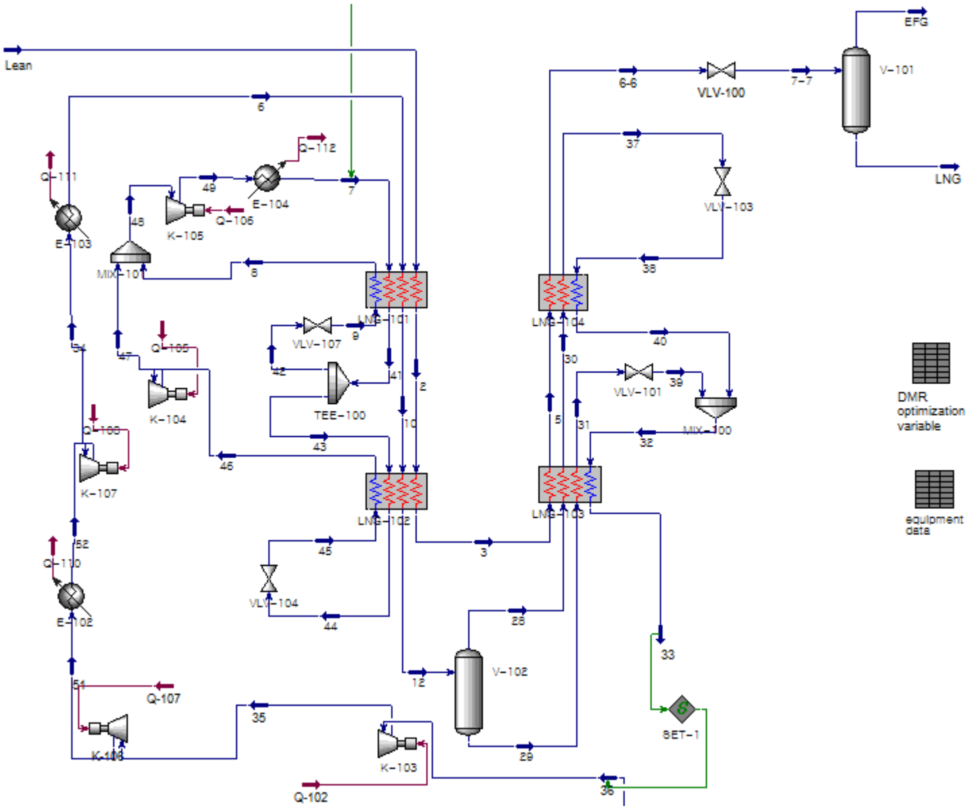

Figure 2-20. DMR base process

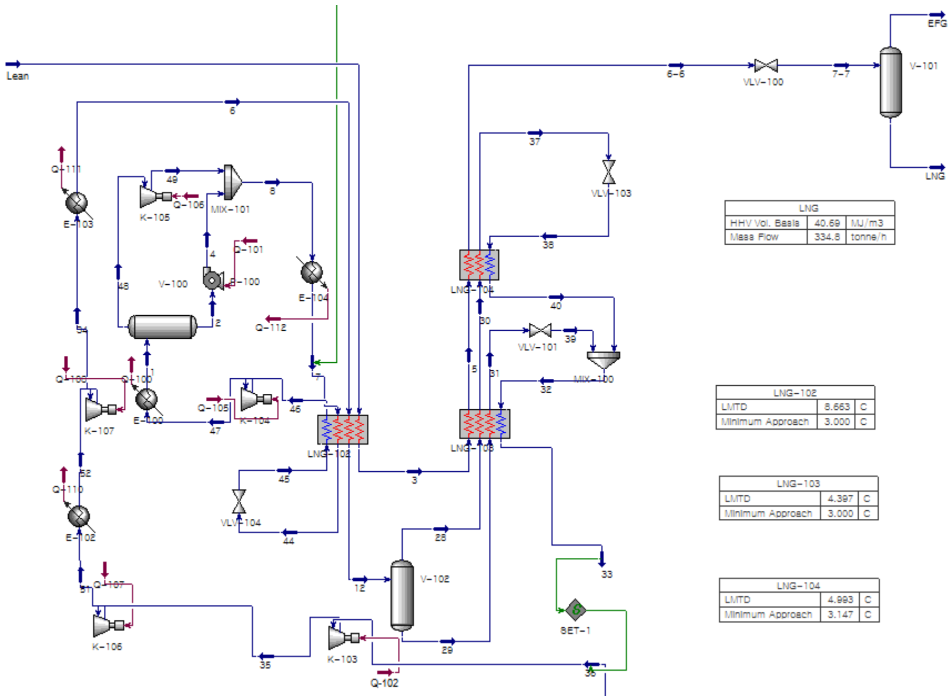

Figure 2-21 DMR process 2

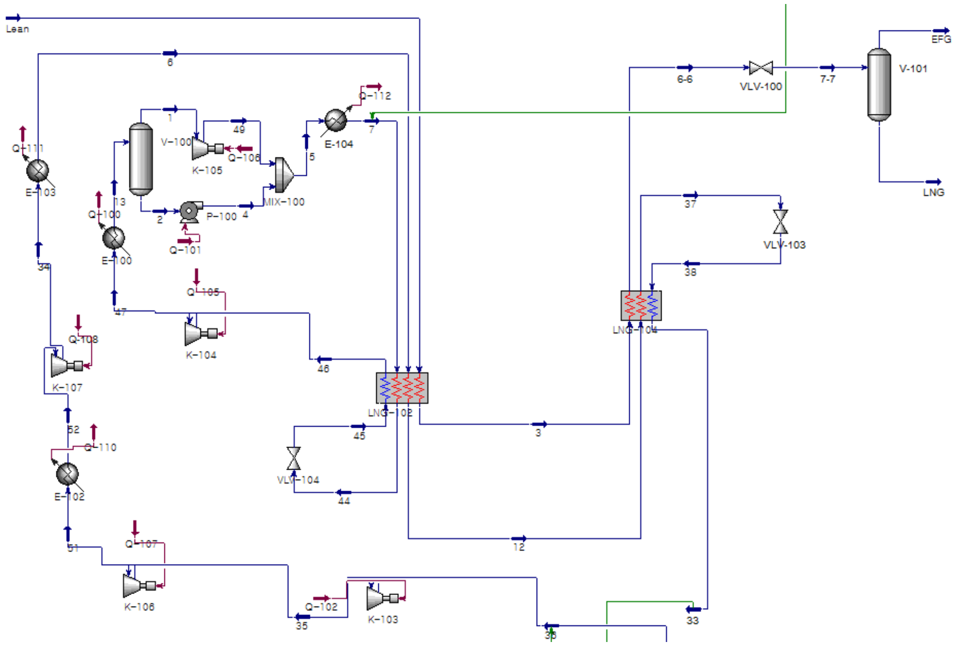

Figure 2-22. DMR process 3

Table 2-6. Liquefaction efficiencies from the previous studies

Author	Journal	Process	Efficiency (kWh/ton)
Barclay & Denton, 2005	LNG journal	Single N2 expander	907.2
Dubar 1998	US 5,768,912	Single N2 expander	647
		Dual N2 expander	453
Lee et al., 2013	ISOPE	Single N2 expander	794
		Dual N2 expander	538
Lim et al., 2014	Industrial & Engineering Chemistry Research	Single N2 expander	679
Shirazi & Mowla, 2010	Energy	SMR	303.6
Khan et al., 2012	Asia-Pacific Journal of Chemical Engineering	SMR	424.4
Moein et al., 2015	Journal of Natural Gas Science and Engineering	SMR	275.04
Lim et al., 2014	Industrial & Engineering Chemistry Research	SMR	347
Vatani et al., 2014	International Journal of Energy Research	SMR	305
Vatani et al., 2014	International Journal of Energy Research	DMR	275

Barclay and Shukri, 2000	Annual Gas Processors Association Convention	DMR	307.2
Venkatarathnam & Timmerhaus, 2008	Springer	DMR	240
Lee et al., 2011	ISOPE	DMR	237.4

2.2.3. Results and discussion

Optimization and economic analysis results of the abovementioned six types of liquefaction processes are shown as below.

- **Optimization results**

Process optimization was performed by minimizing power consumption of the liquefaction cycle as an objective function and the GA, described more detail in section 2.1, connected with HYSYS variables was used in here. Minimizing the required power consumption is commonly used as the objective function of a liquefaction process optimization [72], [65], [68]. There also exist some studies were used exergy efficiency [60], exchangers heat duty [73], heat exchanger area [74], or operating cost as an objective function [75]. Additionally, Minimum temperature approach (MTA) of 3 C was used as optimization constraints in this study.

The results of optimization variables for dual N2 expander process (Figure 2-17) are illustrated as Table 2-7. There are total seven optimization variables including four pressure level for the N2 refrigerant, two inlet temperature for the turbines and one N2 refrigerant split ratio variables. The lowest pressure of N2 refrigeration was 3.92 bar, and the highest pressure was up to nearly 60 bar. The two inlet temperature for the turbines were -17 C and -57.5 C, respectively. N2 refrigerant split ratio was about 0.33. The refrigerant split into dual streams have advantages in the heat integration compared to the single N2 expander process. Accordingly, the liquefaction efficiency could obtain is higher than single N2 expander process. Total power consumption for the liquefaction cycle was 263,092 KW and LNG product was 334.8 ton/h for the Dual N2 liquefaction process. The liquefaction efficiency was 32.75 KW/ton/day (about 785.9 KWh/ton) after optimization, which value was in the range of the previous reported N2 expander efficiency in agreement with Table 2-6.

The optimization variables for SMR process 1 (Figure 2-18) and SMR process 2 (Figure 2-19) SMR liquefaction processes are shown as Table 2-8 and Table 2-9. There are nine and ten optimization variables for the SMR process 1 and SMR process 2, respectively. The SMR process 1 has two-stages compression while the SMR process 2 owns three-stage compression. Hence, the SMR process 2 has one more variable than the SMR process 1. Both processes use same component nitrogen (N2), methane (C1), ethane(C2), propane(C3) n-butane(nC4) as the MR composition variables and one MR mass flow as a variable. The lowest MR pressure for the SMR 1 and SMR 2 processes were 2.76 bar and 3.04 bar. Likewise, the highest MR pressure for the two schemes were 29.59 bar and 43.95 bar. The lowest pressure for the two schemes were similar, but the highest pressure for SMR process 2 was 48.5% higher than SMR process 1. It is because SMR process 2 utilizes three-stage compression systems instead of two-stage compression. The MR mass flow of SMR 1 was 17.4% higher than that of SMR process 2. Total required power and LNG product for the SMR process 1 were 101,170 KW and 334.6 ton/h. For the SMR process 2, the values were 96,322 KW and 334.8 ton/h after optimization. The liquefaction efficiencies of the SMR process 1 and SMR process 2 cycles were 12.6 KW/ton/day (302.4 KWh/ton) and 11.99 KW/ton/day (287.7 KWh/ton), respectively. The liquefaction efficiency of SMR process 2 is about 5% higher than the SMR process 1 after optimization.

The results of optimization variables for the three types of DMR processes (Figure 2-20~2-22) are given as Table 2-10. There are 19 variables for the DMR base process, ten of which are in the warm mixed refrigerant cycle (WMR) and nine in the cold mixed refrigerant cycle (CMR). The WMR cycle is for natural gas precooling which includes three pressure variables (two-stage compression), five composition variables (N2, C1, C2, C3, nC4), one WMR flow ratio and one mass flow variables. The CMR cycle is for liquefaction and sub-cooling containing four pressure variables (three-stage compression), four CMR composition variables (N2, C1, C2, C3) and one mass flow variable. The DMR process 2 and 3 are utilizing only one multi-stream heat exchanger for pre-cooling part. Therefore, they do not include the WMR split-flow ratio variable comparing with the DMR base process, and accordingly both DMR process 2 and 3 consist of total 18 optimization variable. According to Table 2-10 data, the lowest and highest MR pressure for the three schemes show similar values after optimization. The WMR and CMR compositions also have a similar tendency. Total power consumption for the DMR base case, DMR process 2 and 3 were 79,489 KW, 86,410 KW and 89,778 KW respectively. LNG product for those schemes indicate almost the same value (about 334.8 ton/h). The liquefaction efficiency of DMR base was 9.89 KW/ton/day (237.39 KWh/ton), the efficiency of DMR process 2 and 3 were 10.75 KW/ton/day (258.09 KWh/ton) and 11.17 KW/ton/day (268.18 KWh/ton), respectively.

Table 2-11 shows the summary of LNG product and liquefaction performance for each process. The LNG product for each process is similar (at the most about 0.1% difference). The required power consumption and liquefaction efficiency have same tendency. The simplest process N2 expander has highest power consumption and efficiency. By contrast, the most complex process DMR base shows the lowest power consumption and efficiency. The dual N2 expander presents about 69.8% low process efficiency compared to the DMR base scheme. It indicates that the process has more complex configuration has higher process efficiency. DMR base process also shows 21.5%, 17.5%, 8.02% and 11.48% higher process efficiency than SMR process 1, SMR process 2, DMR 2 and DMR 3 schemes.

Table 2-7. Optimization variables of the Dual N2 expander process

Variables		
Pressure (bar)	P_low	3.92
Pressure (bar)	P_m	13.00
Pressure (bar)	P_m2	28.25
Pressure (bar)	P_high	59.37
Turbine inlet Temperature 1 (C)	T_low	-57.51
Turbine inlet Temperature 2 (C)	T_high	-17.06
Refrigerant split ratio	F	0.33
Total number of variables	7	

Table 2-8. Optimization variables of the SMR process 1

Variables		
Pressure	P_low	2.76
Pressure	P_m	10.04
Pressure	P_high	29.59
N2 Composition	C_{N2}	0.07
C1 Composition	C_{C1}	0.27
C2 Composition	C_{C2}	0.34
C3 Composition	C_{C3}	0.14
nC4 Composition	C_{C4}	0.18
MR mass flow	F	1651.00
Total number of variables	9	

Table 2-9. Optimization variables of the SMR process 2

Variables		
Pressure	P_low	3.04
Pressure	P_m	7.15
Pressure	P_m2	21.61
Pressure	P_high	43.95
N2 Composition	C_{N2}	0.09
C1 Composition	C_{C1}	0.26
C2 Composition	C_{C2}	0.33
C3 Composition	C_{C3}	0.06
nC4 Composition	C_{C4}	0.26
MR mass flow	F	1406.03
Total number of variables	10	

Table 2-10. Optimization variables of the DMR processes

Variables			DMR base	DMR 2	DMR 3
WMR cycle	Pressure	P_{PMR-1}	2.80	3.08	3.03
	Pressure	P_{PMR-2}	7.79	7.93	7.56
	Pressure	P_{PMR-3}	16.43	16.34	15.02
	N2 Composition	C_{N2}	0.00	0.00	0.00
	C1 Composition	C_{C1}	0.00	0.00	0.00
	C2 Composition	C_{C2}	0.25	0.22	0.21
	C3 Composition	C_{C3}	0.58	0.58	0.58
	nC4 Composition	C_{nC4}	0.16	0.19	0.20
	Tee Flow ratio	R_{PMR-5}	0.64	-	-
	WMR mass flow	Fwmr	1144.63	970.01	1008.52
CMR cycle	Pressure	P_{CMR-1}	2.91	2.74	2.80
	Pressure	P_{CMR-2}	15.00	14.99	15.00
	Pressure	P_{CMR-3}	20.85	20.44	21.46
	Pressure	P_{CMR-4}	45.99	43.79	50.33
	N2 Composition	C_{N2}	0.06	0.05	0.11
	C1 Composition	C_{C1}	0.41	0.40	0.36
	C2 Composition	C_{C2}	0.34	0.33	0.32
	C3 Composition	C_{C3}	0.20	0.21	0.22
	CMR mass flow	Fcmr	690.88	688.28	709.85
Total number of variables		19	19	18	18

Table 2-11. LNG product and liquefaction performance by each scheme.

	N2 Dual expander	SMR 1	SMR 2	DMR base	DMR2	DMR3
LNG Product (ton/h)	334.8	334.6	334.8	334.8	334.8	334.8
Power consumption (kw)	263,092	101,170	96,322	79,489	86,410	89,778
Liquefaction efficiency (kw/ton/day)	32.75	12.60	11.99	9.89	10.75	11.17

- **Economic evaluation**

The capital cost and operating cost estimation method are the same as that described in Chapter 2.1. The results by each scheme are shown as Figure 2-23.

The SMR process 2 gives the lowest total capital cost among the processes. It was 5.13%, 9.77%, 14.09%, 8.19% and 9.62% lower than the N2 expander, SMR process 1, DMR base, DMR 2 and DMR 3 processes, respectively. N2 expander process has simpler configuration than the SMR process 2 but shows a little higher capital investment cost. This is because the N2 expander process uses expensive turbines unlike the SMR process 2 using JT valve. Similarly, the SMR process 1 presents nearly 10 % higher capital expenditure compared with the SMR process 2 even though it owns simpler process complexity. Because SMR process 1 requires higher compressor size than SMR process 2 which causes great effect on the total capital expenditure. DMR base process, which has the most complex configurations in here, indicates about 14% higher capital expenditure compared to the SMR process 2.

The total operating cost by each scheme (Figure 2-23) shows that DMR base process reveals the lowest operating cost per year. Conversely, due to a poor liquefaction efficiency the N2 expander process presents the highest operating cost which is 54% higher compared to the DMR base scheme. Operating cost savings of DMR base case were 21.43% and 17.48% compared to SMR process 1 and SMR process 2. For the other DMR processes: DMR process 2 and 3 the savings were 8.01% and 11.46%, respectively. Even though DMR base scheme gives the highest total capital cost expenditure among the schemes, the operating cost shows lowest expenditure due to the higher process efficiency than the other processes. For example, DMR base process presents about 14% higher total capital cost compared to SMR process 2. On the other hand, it spends around 17% lower operating cost than the SMR process 2 per year.

Total annualized cost (TAC) in this chapter is defined as the sum of annualized TCC and total operating cost as Equation (2-8). The annualized TCC is calculated based on an annual basis in terms of an interest rate of 5% over 10-year period for economic evaluation. TAC has the advantage that reflect both total capital cost and operating cost impact.

$$TAC = Annualized\, TCC + Total\, operating\, cost \qquad (2\text{-}8)$$

Figure 2-24 presents TAC expenditure for each scheme. According to the figure, DMR base process shows the lowest TAC value while the N2 expander process shows the highest. The difference is around 50.92% between the two processes. TAC of DMR base process reveals 18.85% and 14.33% lower than the SMR process 1 and SMR process 2 It also shows 6.37% and 9.66% lower compared to the DMR process 2 and 3. The TAC and total operating cost for each process tend to be similar, but the rate of TAC compared to total operating cost was slightly decreased due to the effect of total capital cost. However, it is easy to find that the cost impact of operating cost is much greater than that of total capital cost.

Profitability results for the process schemes are given as Table 2-12. LNG price in this study was assumed 8 USD/MMBTU. The product revenues for the six different processes are similar, with a difference of less than 0.1%. It means that all the processes give similar LNG production. The gross profit of DMR base scheme shows the highest while the N2 expander process presents the lowest value among the schemes. The difference in gross profit between these two processes is approximately 14.05%. The profit of SMR process 2 shows 2.26% lower than the DMR base process. However, the SMR process 2 gives the shortest payback time and DMR base process shows the longest (the deviation is approximately13.84% between the two processes).

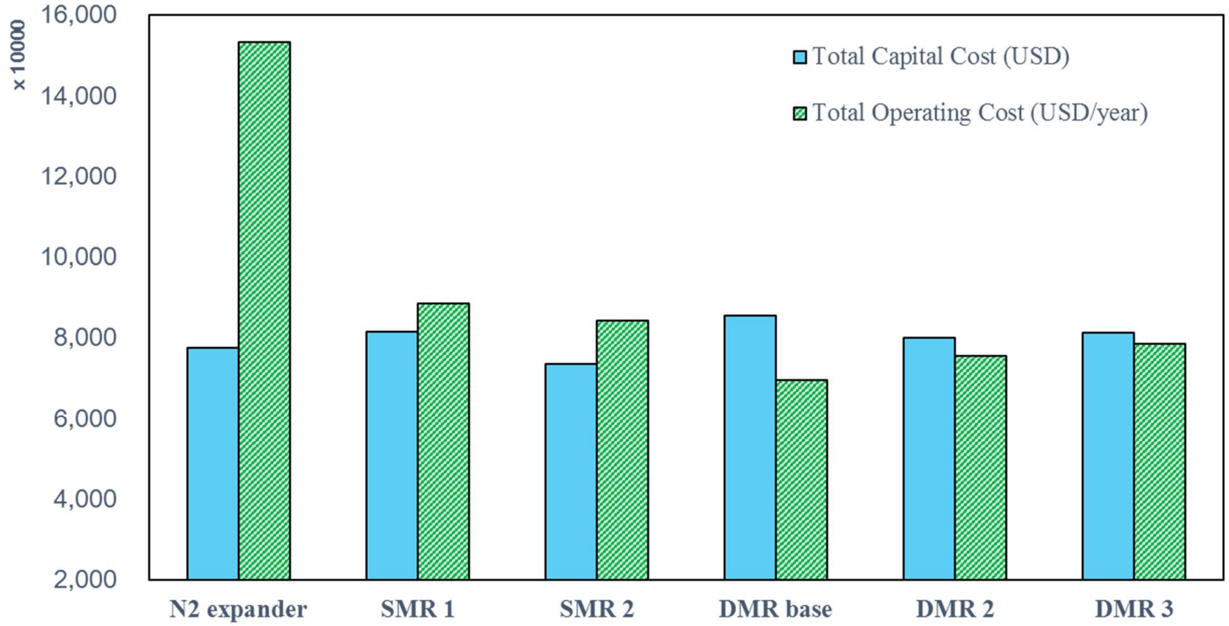

Figure 2-23. Total capital cost and operating cost for each scheme

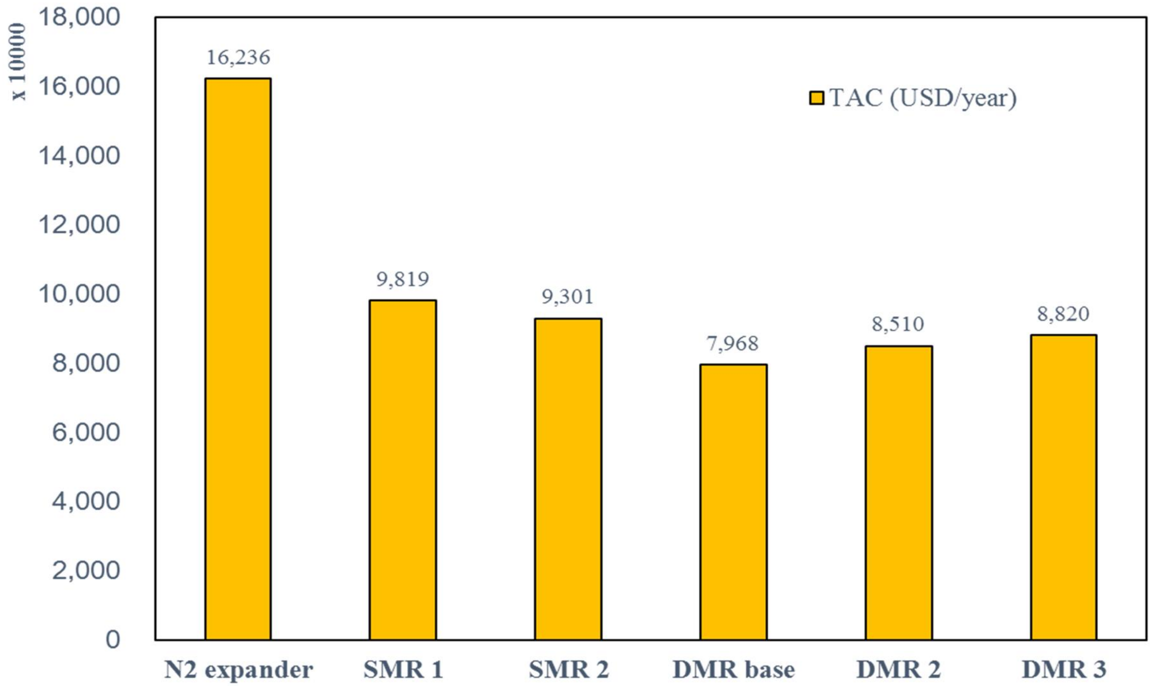

Figure 2-24. Total annualized cost (TAC) for each scheme

Table 2-12. Profitability analysis for the process schemes.

	N2 Dual expander	SMR 1	SMR 2	DMR base	DMR 2	DMR 3
Product revenue (USD/year)	1,215,912,298	1,215,202,998	1,215,912,298	1,216,211,499	1,216,061,919	1,215,912,298
Gross Profit (USD/year)	597,204,635	661,129,017	666,084,179	681,129,916	674,917,345	671,817,675
Net Profit (USD/year)	418,043,244	462,790,312	466,258,925	476,790,941	472,442,141	470,272,373
Payout time (year)	0.185	0.176	0.158	0.179	0.169	0.173

In this chapter 2.2, process performance and economic evaluation were carried out on six types of representative liquefaction processes for offshore application. The results show that dual N2 process requires highest power consumption (more than 60% compared to DMR base) and lowest net profit (approximately 10% lower than the other schemes) among the schemes. Therefore, it can be seen that the N2 expander process is not advantageous in terms of process performance and economy aspects compared with other processes. The SMR 2 process shows the lowest total capital cost and payout time while DMR base case presents the highest net profit and lowest operating cost among the processes. The SMR 2 shows 14% and 12% lower total capital cost and payout time than DMR base. Conversely, DMR base process gives lower total operating cost compared to the SMR 2. Both the SMR 2 and DMR base each have advantages, compared with other processes, that should be considered as good candidate schemes for offshore units.

2.3. Evaluation of a representative NGL/LNG integrated process considering LNG HHV specification

2.3.1. Overview

In a conventional normal or rich feed condition, integrate the NGL recovery and liquefaction process have advantages over economic performance because integrating the two series cryogenic processes can share refrigeration systems and also eliminate a number of process equipment. Cueller et al. 2002 presented that integrated NGL/LNG co-production design could incredibly reduce capital and operating costs [76]. NGL are always removed from original treated feed gas for several reasons. For example, the added value of heavier hydrocarbons, the pipeline specification requirement for transportation etc. Moreover, the heavier hydrocarbons such as C5+ should be removed prior to liquefaction process in order to prevent freezing during liquefaction. Cryogenic processes are normally the most economic for the NGL recovery which have shown by previous studies and industrial experience. The representative processes are ISS, GSP and IPSI process etc. They use expander and joule-thomson (JT) valve to letdown the feed pressure for achieving cryogenic condition, these processes sometimes also need additional external refrigeration systems when the feed composition is rich. There are several existing liquefaction processes licensed by different companies such as APCI, Shell, Linde etc. The representative schemes are APCI SMR, C3MR and DMR processes, among them C3MR process has the dominant market share, especially in onshore plant, so far. However, the DMR process ,which is replaced the propane precooling cycle instead of the mixed refrigeration cycle, has reported its larger train capacity and higher liquefaction efficiency than C3MR [70],[77]. Both the NGL recovery process and liquefaction process are need refregeration systems, so they may have the potential to integrate the two process configurations. Elliot et al. 2005 applied an integrated concept to real NGL/LNG projects and they showed that proper integration results in significant deacrease of overall capital cost and improving products production as well [78].

Pevious studies could classify as two groups regarding the NGL and LNG integrated processes. First, the researches focus on developing process efficiency. Ghorbani et al. 2016 introduced an integrated process for LNG and NGL products including a nitrogen rejection unit, the process applied C3MR cycle for providing refrigeration and the results showed resonalble specific power with more than 90% NGL recovery [79]. Mehrpooya et al. 2014 proposed three integrated processes with applying MFC, DMR and C3MR refrigeration systems, they reported that these configurations had high ethane recovery and lower specific power compared to the similar previous studies [80]. An integrated NGL/LNG process configuration was examined by Vatani et al., which could be applied for large scale LNG plants. This scheme utilized dual mixed refrigeration cycles for liquefaction, which showed good efficiency and acceptable NGL recovery under a rich feed condition [22]. Khan et al. 2014 presented integrated process using SMR cylcle and energy intensive coupled distillation configurations. After knowledge-based optimization, the proposed scheme showed remarkable improvement in compression power saving compared to the base case [81]. Wang and Xu 2014 presented that an integrated NGL recovery with LNG re-gasification process showed remarkable potentials on both energy savings and product production [24]. Hudson et al. 2003 reported several examples and showed that a good recovery rate as well as efficiency performance could be achieved by integrating liquids recovery with liquefaction processes [82]. Brostow and Roberts 2013 indicated that the required energy could save remarkbly by integration of NGL recovery and liquefaction processes [31]. Dynamic simulation was performed by Husnil et al. 2014 in order to determine the control variables for a modified SMR cycle integrated with NGL recovery process [83]. Pillarella et al. 2007 introduced an integrated NGL/LPG extraction, a scrub column was applied, with C3MR liquefaction cycle. This simple integration process concept required less equipment than traditional extraction before liquefaction processes [84]. Uwitonze et al. 2016 compared DMR liquefaction cycle integrated with a conventional NGL recovery process and the other two proposed cases, which improved NGL recovery section with heat integration by distillation column [85]. They investigated that the heat integration process showed lower overall energy consumption and the products purities could be acquired by the integrated column system.

Second, the previous studies consider both efficiency and economy analysis. He and Ju 2014 proposed an integrated NGL recovery scheme with a signgle mixed refrigerant cycle, they selected global optimization method of GA for optimizing the process and described that the energy consumption could be reduced more than 9% as well as the economic analysis showed good profitability [86]. Lee et al. 2012 investigated and compared natural gas liquefaction and recovery processes for offhsore application, they used SMR cycle as a refrigeration system due to its simple, compact and suitable characteristic on offshore floating structures [87]. Exergoeconomic analysis was performed by Ghorbani et al. 2017 for a NGL recovery and liquefaction integrated process, the results showed that exergy efficiency by air coolers and exergy destruction in the compressors affect power consumption straightly [88]. Ansarinasab and Mehrpooya 2017 evaluated two integrated processes using DMR and MFC refrigeration cycles by adopting advanced exergoeconomic analysis, the results showed that the heat exchanger plays a crucial role in the coproduction of NGL and LNG products process [89]. Ansarinasab et al. 2016 performed exergoeoconomic evaluation on a recently reported integrated process which used MFC by refrigeration system, the results showed that the most important factor for the excergy destruction is due to heat exchangers [90]. A novel NGL/LNG integrated process using absorption system in precooling and MFC cycle for liquefaction was proposed [91]. The exergy analysis showed 12.72% improvement for overall efficiency compared to base case due to adopting absorption precooling system and the proposed process could increase 6.2% of net annual benefit according to economic analysis results.

2.3.2. A conventional integrated process description

Figure 2-25 illustrates a NGL/LNG conventional integrated process flow diagram. The pretreated feed, without water and acid gas, stream at 30 °C and 60 bar is sent to the E1 and E2 multi stream heat exchangers successively for precooling the feed stream. Then the precooled stream 2 is flashed via V1 separator. The vapor stream is divided by stream 3 and stream 4, where the stream 3 is further cooled by the column top stream 9 and enters column top stage after depressurizing to the column pressure through JT valve. The other stream 4 is directly depressurized by the turbo expander C1 before introducing to the column. The liquid stream 5 is first expanded by JT valve to the column pressure and then enters to the column. The column top stream 9, after heat exchanging with stream 3, is compressed about 60 bar by C2 and C3 compressors before feeding to the E3 multi stream exchanger. The stream 13 follows to the exchanger E4 for sub cooling about -160 °C, then the stream 14 is reduced its pressure to about 1bar by JT valve. After

JT expansion the stream may contain small amount of end flash gas which could separate from a separator. The remaining liquid product is sent to LNG tank for storage.

It can be divided into two cycles, precooling cycle and liquefaction cycle, for the DMR liquefaction part. The PMR (precooling mixed refrigerant) cycle is the inner loop cycle with the multi stream heat exchangers E1 and E2. The PMR-3 stream, which is compressed about 16.7 bar and the air cooled temperature of 33 °C, enters the first multi stream exchanger E1 and cooled about 0°C. The PMR-4 stream is divided two streams; one portion is expanded by JT valve to about 7.83 bar and returned to the E1 exchanger for providing cold refrigeration, the other remaining portion is further cooled through exchanger E2 about -33°C and then depressurized by JT valve about 2.81 bar before returning to the E2 exchanger for providing cold refrigeration. The CMR cycle for liquefaction is the outer loop cycle extended with E1, E2, E3 and E4 multi stream heat exchangers. The pressurized CMR-4 stream is, about 46 bar, first cooled to 33°C via the air cooler and follows to the E1 and E2 exchanger successively. After precooling section, the CMR-6 stream is cooled about -33°C and then flashed to the CMR-7 and CMR-8 stream through a cold separator. The vapor stream CMR-7 is introduced to the E3 and E4 exchangers for further cooling and the pressure reduced to about 2.9 bar after expansion via JT valve. Then the stream cold stream is returned to the E4 exchanger for providing cold refrigeration. The liquid stream CMR-8 is passed the exchanger E3 and depressurized about 2.9 bar by JT valve and mixed with the returning cold stream from the E4 heat exchanger. The mixed cold stream follows the exchanger E3 in order to provide cold refrigeration. Then the stream CMR-1 increase the pressure via three stage compressors for complete the cycle.

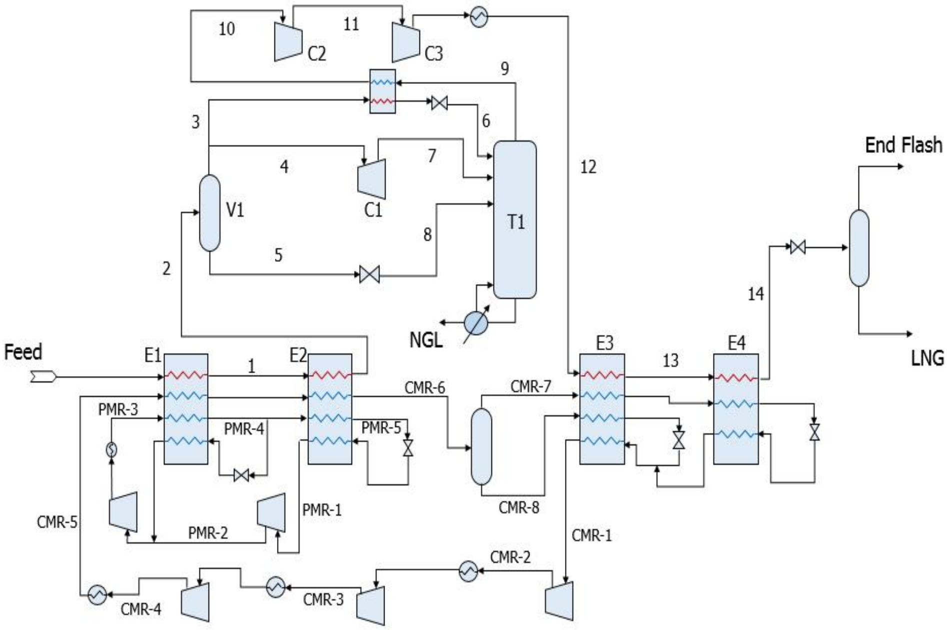

Figure 2-25. The conventional NGL/LNG integrated process (base case)

CHAPTER 3. The proposed integrated processes under the lean feed conditions*

3.1 Overview

According to the chapter 2.3.1, previous studies for the integrated NGL recovery and liquefaction schemes, some of them have tended to only focus on improving process efficiency. For the large scale plants, plants efficiency maybe one of the most important factors compared to the total capital investments in terms of long-term consideration. On the contrary, not only proper process efficiency but also the capital investments should be considered for small-medium scale plants (offshore units etc) or peak shaving plants. There are also some papers evaluated both process efficiency and economy evaluation. For example, He and Ju 2014 proposed a novel integrated process and performed exergy analysis and profitability analysis as well [86]. Lee et al. 2012 designed the integrated processes which could applicable for offshore units with consideration of investment costs [87]. Nevertheless, the studies tended to concentrate on a conventional normal rich feed condition and also without considering the LNG higher heating value (HHV) specification. The feed compostion is a very important factor in the natural gas processes [8]. Different feed conditions may show completely different process performance. Moreover, with the increasing number of unconventional lean gas reservoirs also require a re-evaluation of the integrated process configurations under the lean feed compositions. Additionally, the minimum requirements of LNG HHV in East Asia such as Japan and South Korea are higher than Europe countries or U.S. The two countries market share in LNG market is reported more than 47%

* *This chapter references the author's accepted journal paper: C. Jin, Y. Lim, Optimization and economic evaluation of integrated natural gas liquids (NGL) and liquefied natural gas (LNG) processing for lean feed gas, Applied Thermal Engineering.*

based on IGU report [1]. This implies that the HHV specification for LNG products is also an important factor which should be included for consideration in the LNG processes.

In this paper we analyzed both process efficiency and economic evaluation with consideration of LNG HHV specification under lean feed conditions. We evaluated specific power consumption for liquefaction process and total annualized costs of selected integrated schemes which were conventional integrated NGL recovery with liquefaction process and proposed two cases of simplified integrated processes, respectively.

3.2 The proposed integrated process description

3.2.1. The proposed integrated process with DMR cycle

The simplified integrated process scheme utilizing DMR liquefaction cycle is shown in Figure 3-1. The simplified process has different configurations in NGL recovery section compared to Figure 2-24 base case. The previous study conducted by Jin and Lim shows the NGL recovery scheme could be simplified under lean feed condition.[92] The pretreated feed at 30°C and 60 bar follows the multi stream exchangers E1 and E2 successively for precooling the feed temperature of -33°C. The stream 2 is directly letdown the pressure to the column pressure through JT valve and introduced to the T1 column top stage. After extracting NGL from fractionation, the column top gas stream is compressed about 60 bar via C1 compressor and fed to the E3 and E4 exchangers, respectively. The stream 8 reaches the temperature of -160 °C after expands through JT valve to slightly above atmosphere pressure for stripping end flash gas. After that, the liquid LNG product send to LNG tank for storage. For the DMR liquefaction process, PFD is same as the base case and do not mention here again.

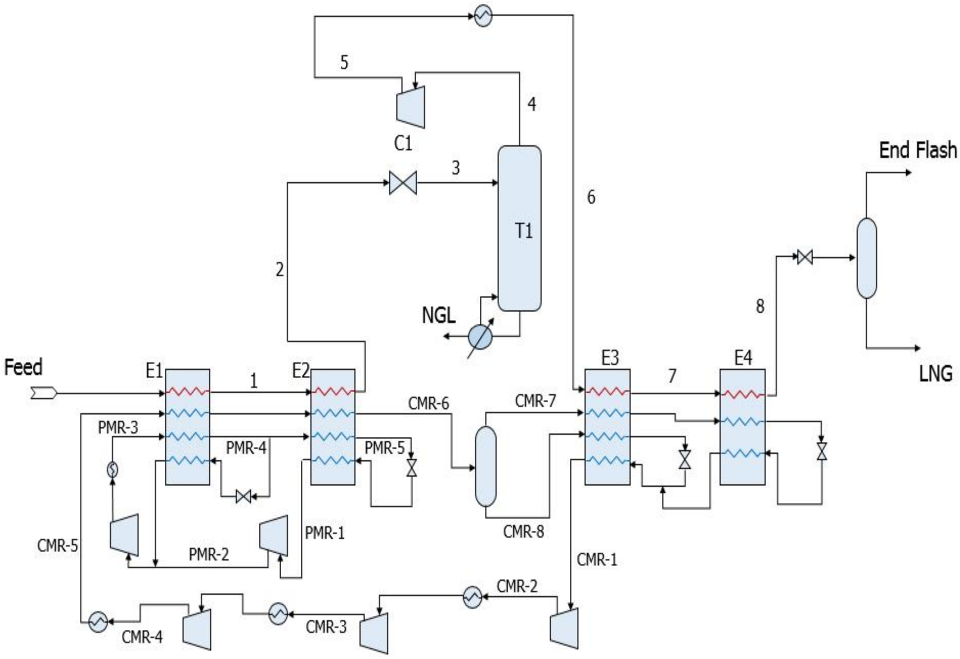

Figure 3-1. The proposed NGL/ LNG integrated process (case 1)

3.2.2. The proposed integrated process with SMR cycle

The simplified integrated process scheme adopting SMR liquefaction cycle is shown in Figure 3-2. This simplified integrated process with SMR liquefaction system is the same configuration in NGL recovery part as the Figure 3-1 scheme. The pretreated feed follows the multi stream heat exchangers E1 for precooling the feed temperature about -33°C. The stream 1 is reduce the pressure by JT valve before entering the T1 column top stage. The column top gas stream 3 is compressed about 60 bar thorough C1 compressor and fed to the E2 exchanger.

The SMR cycle has only one mixed refrigerant cycle. Therefore, it has more simple configuration than the previous two cases (Figure 2-24 and Figure 3-1). The MR-1 stream is elevated pressure about 42 bar by the C2 and C3 compressors. The MR-3 stream is cooled to about 30C through air cooler system, then it is separated by vapor and liquid streams through a separator. The vapor stream MR-4 is further compressed by a compressor and the liquid stream MR-5 increase the pressure by a pump before mixing each other. The MR-6 stream is divided by vapor stream MR-7 and liquid stream MR-8 via a separator. The stream MR-7 and MR-8 are precooled to -30°C after passing the first E1 exchanger. Then both the MR-9 and MR-10 streams pass the E2 heat exchanger for further cooling. After JT expansion the cold stream MR-11 and MR-12 returned to the E2 and E1 exchangers sequentially for providing cold refrigeration.

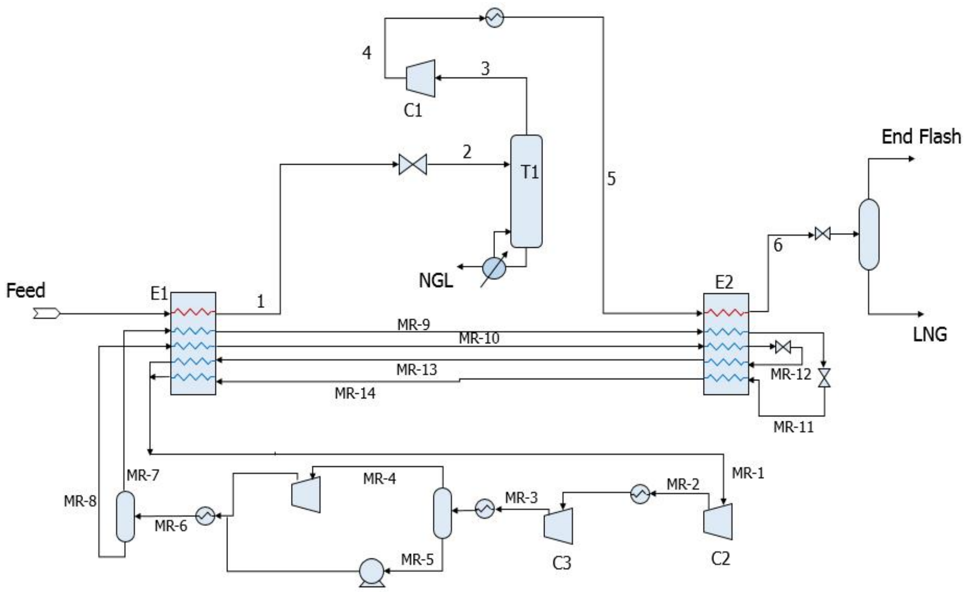

Figure 3-2. The proposed NGL/ LNG integrated process (case 2)

3.3 Feed compositions and specification

Feed composition is a decisive factor in process selection. For a same process with different feed composition, the process configuration may be changed for the process performance. We select a typical lean feed composition as shown in Table 3-1 since we focus on the lean feed condition, where the GPM value of this feed is 1.84.[92] The GPM is the amount of recoverable liquid which is expressed in gallons per 1000 standard cubic feet of gas at 60 °F and basically consider as lean feed when the value is lower than 2.5 [32]. The feed composition in Table 3-1 is not include water and acid gas components because the NGL recovery and liquefaction processes are the post process of AGRU and dehydration processes.

Specify common key parameters are very important for fair comparison of different process schemes because these parameters could affect process performance significantly. Table 3-2 shows the common process specification in this study. The minimum HHV is specified in order to satisfy the LNG specification of east Asian countries. The LNG C5+ specification is referenced from gas processors suppliers association (GPSA) engineering data book [33].

Table 3-1. Feed composition (mol. %)

Component	C1	C2	C3	iC4	C4	iC5	C5	C6+	N2
Feed	92.97	4.18	1.23	0.54	0.54	0.04	0.04	0.03	0.43

Table 3-2. Common process key specifications

Feed flow rate	19,519 kg mole/h
Plant inlet gas pressure	60 bar
Plant inlet gas temperature	30 °C
Compressor/Turbine/Pump efficiency	80%
Pressure drop across the heat exchanger	0.5 bar
Heat exchangers minimum temperature approach	3 °C
HHV (LNG)	Min. 40.50 MJ/Sm^3
C_{5+} (LNG)	0.1 mol% max.
Property fluid package	Peng Robinson

3.4 Liquefaction efficiency analysis

In a liquefaction process the efficiency is normally expressed as the consumed power of refrigeration cycle compressors over produced LNG. Therefore, the minimization of the total required work of compressors stands for better liquefaction efficiencies. In this study, the objective function is expressed as Equation (3-1), minimization the total work of refrigeration cycle compressors [93], [94], [65].

$$Minimize\ f(X) = \sum W_{Compressors} \qquad (3\text{-}1)$$

In Equation (4-1) the X is optimization variables including MR mass flow of each component, the outlet pressure of compressors etc. The optimization variables (X) in this study is shown as Table 3-3 and Table 3-4. The WMR for precooling cycle uses N2, C1, C2, C3, nC4 refrigerants and N2, C1, C2, C3 are selected as the CMR for the DMR liquefaction cycle and N2, C1, C2, C3, nC4 refrigerants are selected for the SMR liquefaction cycle

The constraints of this study are minimum temperature approach in the heat exchangers, compressors inlet stream temperature. These constraints are detailed as follows:

$$\Delta T_{min,HX-i} \geq 3\ °C \ ;$$

$$T_{inlet,compressor-i} \geq T_{dew,i} \qquad (3\text{-}2)$$

Where the $\Delta T_{min,HX-i}$ stands for the MTA in heat exchanger i, $T_{inlet,compressor-i}$ and $T_{dew,i}$ refer to operating temperature of i^{th} inlet compressor stream and dew point temperature of the stream.

Table 3-3. Optimization variables (with DMR cycle)

WMR cycle	Pressure	P_{PMR-1}
	Pressure	P_{PMR-2}
	Pressure	P_{PMR-3}
	N_2 composition	C_{N2}
	C_1 composition	C_{C1}
	C_2 composition	C_{C2}
	C_3 composition	C_{C3}
	nC_4 composition	C_{nC4}
	Tee flow ratio	R_{PMR-5}
	WMR mass flow	F_{WMR}
CMR cycle	Pressure	P_{CMR-1}
	Pressure	P_{CMR-2}
	Pressure	P_{CMR-3}
	Pressure	P_{CMR-4}
	N_2 composition	C_{N2}
	C_1 composition	C_{C1}
	C_2 composition	C_{C2}
	C_3 composition	C_{C3}
	CMR mass flow	F_{CMR}
Total number of variables		19

Table 3-4. Optimization variables (with SMR cycle)

MR cycle	Pressure	P_{low}
	Pressure	P_{m1}
	Pressure	P_{m2}
	Pressure	P_{high}
	N_2 composition	C_{N2}
	C_1 composition	C_{C1}
	C_2 composition	C_{C2}
	C_3 composition	C_{C3}
	nC_4 composition	C_{C4}
	MR mass flow	F
	Total number of variables	10

The GA is used for the process optimization. The GA Matlab code linked with Aspen HYSYS is used for obtaining the optimum process conditions. The GA is a random search method based on the idea of natural evolution which uses historical information and eventually find the global optimum by reproducing a series of population. The GA method is an effective tool for hydrocarbon process optimization and used for several previous studies [86], [95]. The population size 200 and the elite count 10 are selected for the GA main tuning parameters.

3.5 Economic evaluation

Economic evaluation as well as profitability analysis between the integrated base process and the proposed two schemes were performed the same method as described in Chapter 2.1. The TAC in this research is defined the same as Equation (2-8) which is discussed in previous Chapter 2.2. The annualized TCC is calculated by an annual basis and the interest rate of 5% over 10-year period was used in this study.

3.6 Results and discussion

The total capital cost by each scheme is shown in Figure 3-3. The conventional LNG/NGL integrated process (base case) shows the highest total capital cost, whereas the simplified integrated process utilizing SMR cycle (case 2) presents the lowest total capital cost. The total capital cost difference of these two schemes is about 21.5% because case 2 scheme not only simplified NGL recovery section but also liquefaction cycle which could lead to significantly reduce the equipment counts. The integrated process using DMR cycle (case 1) also can save the total capital cost 4.3% compared to the base case due to the simplified NGL recovery section.

Figure 3-4 indicates the total operating cost for the three different processes. The case 1 scheme shows the lowest total operating cost compared with other processes. The saving is 1.5% and 23.9% compared to the base case and case 2 schemes, respectively. The case 2 scheme gives highest operating cost due to the highest compressor power consumption compared with base case and case 1 processes which use DMR instead of SMR cycle for high liquefaction efficiency.

The TAC for the processes is illustrated by Figure 3-5. The case 2 gives the highest and, conversely, the case 1 shows the lowest TAC. The case 1 scheme has 1.8% lower TAC than the base case scheme and 18.5% lower TAC than compared to the case 2. The trend of TAC is similar with the total operating cost as shown in Figure 3-4. It means that the impact of total operating cost is greater than the total capital cost in terms of TAC.

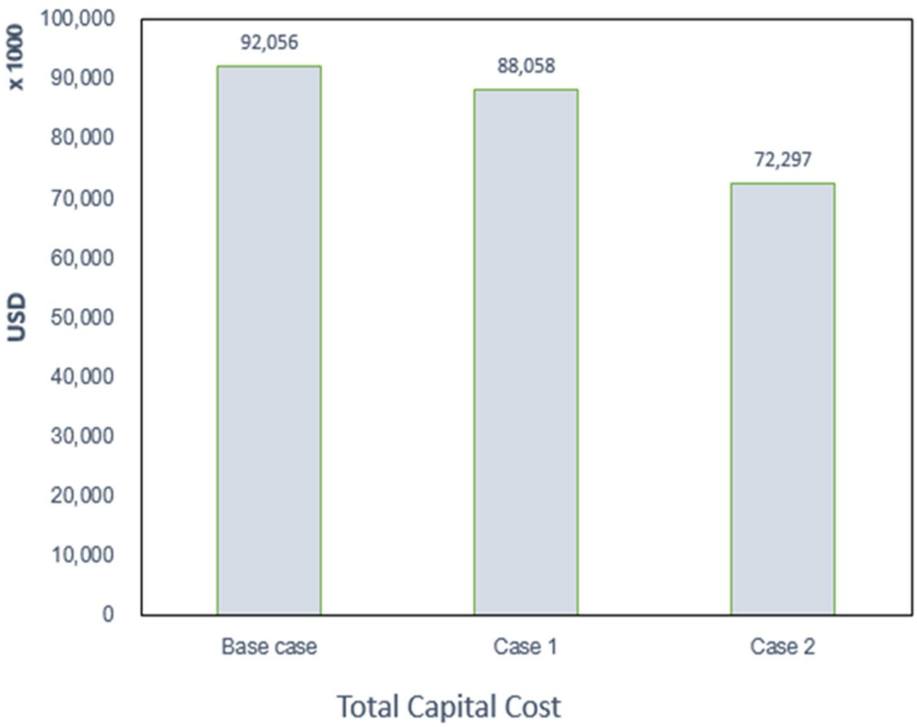

Figure 3-3. Total capital cost for each scheme

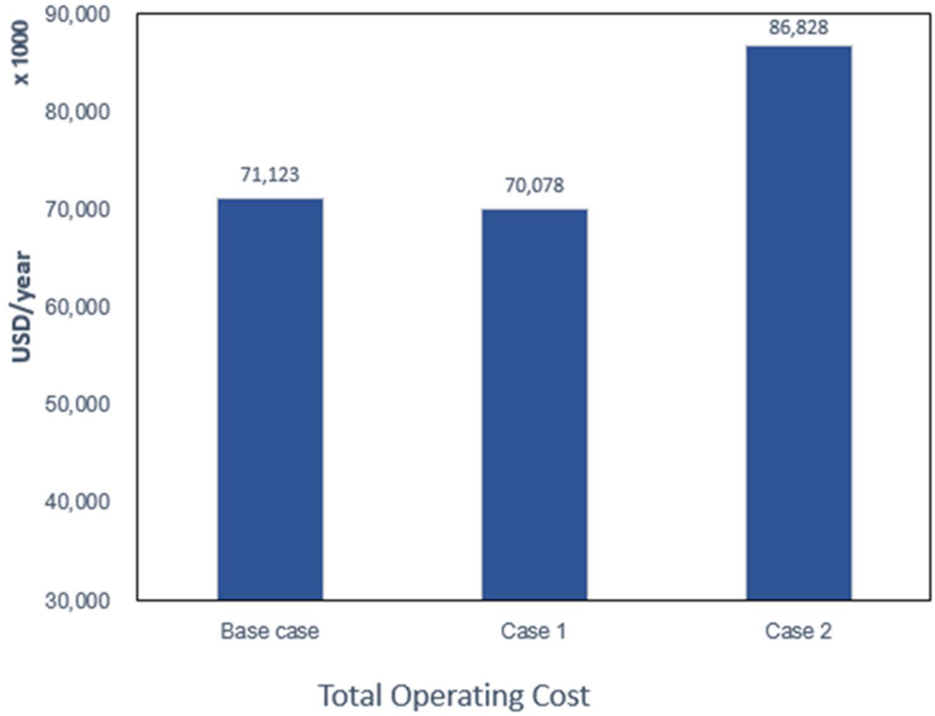

Figure 3-4. Total operating cost for each scheme

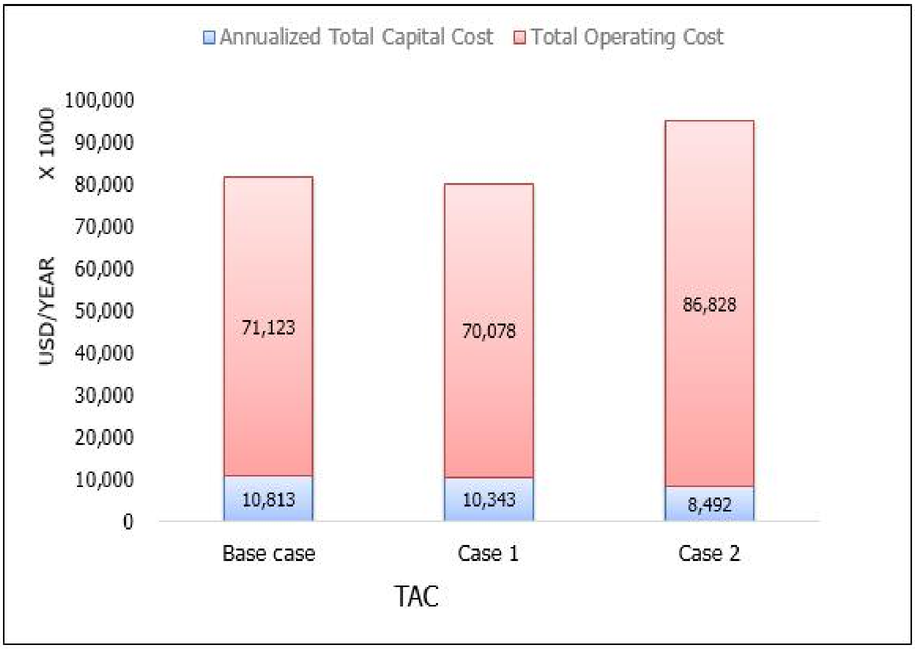

Figure 3-5. TAC for each scheme

The LNG/NGL product and liquefaction efficiency for the candidate schemes are shown in Table 3-5. The LNG product of each process is similar, and there is less than 0.3% deviation between the schemes. The NGL product recovery is very small for all cases because this study we focused on lean feed conditions that has only small amounts of heavy hydrocarbons. The LNG product and consumed compressor power are the two factors that affect the liquefaction efficiency. The base case and case 1 processes have almost the same specific power (efficiency). Case 2 process presents the highest specific power, about 24% higher, compared to the other two schemes. This is because case 2 adopts SMR cycle as the liquefaction part, which has a poor liquefaction efficiency compared with DMR cycle that requires more compressor power than the other two schemes.

A profitability analysis was also carried out for each scheme as presented in Table 3-6. The product revenue for the three schemes are almost the same (less than 0.1%). However, the net profit and gross profit show the same trend for case 1 and the base case, these are 2.52% higher than those of case 2. This is mainly due to the total operating cost difference. The net profit for the base case and case 1 are nearly the same (only 0.01% deviation). The payback time for case 2 process shows the lowest value, even though case 2 process has the lowest net profit. The payout time for case 2 is 19.4% and 15.8% lower than the base case and case 1, respectively. This is because the total capital cost of case 2 is remarkably lower than the other processes as shown in Figure 3-4.

Table 3-5. LNG/NGL product and liquefaction performance

	Base case	Case 1	Case 2
LNG product (ton/h)	333.1	332.3	332.1
NGL product (ton/h)	1.8	2.6	2.8
Compressor power(kW)	79,798	79,758	98,879
Specific power (kW/ton/day)	9.983	10.000	12.407

Table 3-6. Profitability analysis comparisons by each scheme

	Base case	Case 1	Case 2
LNG product revenue($/year)	1,209,709,828	1,206,837,991	1,206,067,943
NGL product revenue ($/year)	5,089,192	6,971,171	7,407,367
Product revenue ($/year)	1,214,799,019	1,213,809,162	1,213,475,309
Gross Profit ($/year)	678,227,230	678,281,762	661,198,337
Net Profit ($/year)	474,759,061	474,797,233	462,838,836
Payout time (year)	0.194	0.185	0.156

The net revenue in this study is defined as net profit multiply by time and minus total capital cost in order to compare the revenue which is both consider net profit and equipment cost impact. The net revenue of each scheme by years is presented in Figure 3-6. According to Figure 7, the net revenue of case 2 scheme is higher than the other two schemes before 1.3 years and after passing 1.3 years the net revenue of case 1 scheme is higher than the other two processes (the years may be longer than 1.3 years in a real plant because the total capital cost calculated by academic methods normally tends to be underestimated compared to the actual practice). It means that if the plant reservoir life is less than 2 years, like peak shaving plants or shale gas reservoirs etc, the case 2 scheme can be consider as one of the candidate processes. However, if the plant reservoir life, as the traditional exploited plants, is longer the case 1 is a good option.

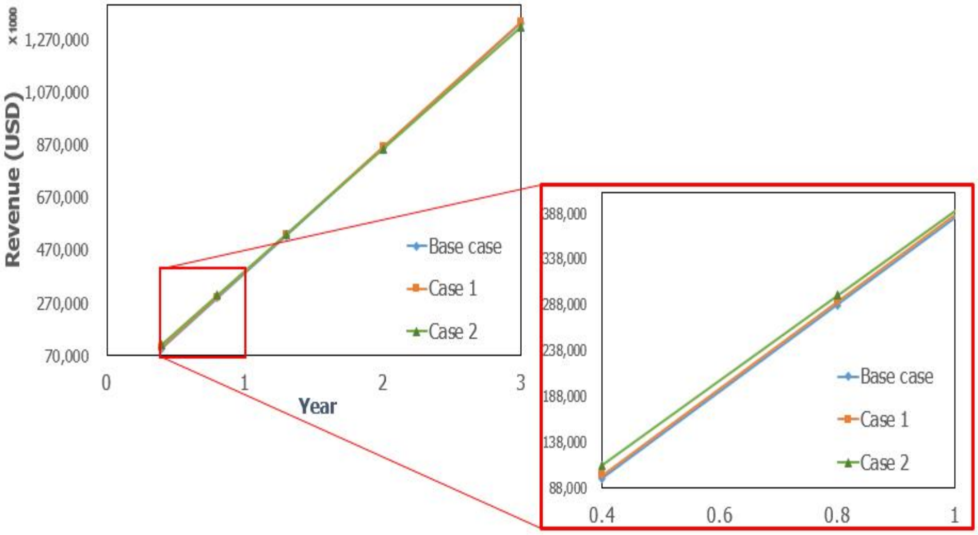

Figure 3-6. Net revenue (net profit excluded capital cost) by time

3.7 Summary

Liquefaction efficiency optimization and economic evaluation are performed for the three cases of LNG/ NGL integrated processes under the lean feed condition. The GA algorithms linked with commercial software HYSYS is used for optimizing the process efficiency considering LNG HHV specification value. The economic evaluation is investigated by each process include the TAC, total capital cost, total operating cost, net profit, payout time and net revenue. The results show that the proposed integrated scheme case 1, which is simplified NGL recovery section compared with the base case, has higher net profit and lower capital cost compared with the conventional base case scheme without notable loss of liquefaction efficiency under lean feed condition. It is the opposite results compared with former related integrated process studies applied under normal or rich feed condition. It means that when the feed composition is considerably lean the NGL recovery section can be simplified without remarkable loss of process efficiency for saving capital cost.

The simplified case 2 scheme, which is simplified both liquefaction and NGL recovery parts compared with the base case, gives about 24% lower liquefaction efficiency compared to the base case and the case 1. Moreover, the case 2 shows highest TAC and lowest net profit among the three schemes. However, it has the lowest total capital cost and payout time because of the less equipment counts. Additionally, the net revenue of case 2 shows higher value than the other two schemes when the plant operation time is less than 1.3 years. Therefore, when the plant reservoir life is short enough, such as some peak shaving plants, shale gas reservoirs, special offshore units and so on, the simplified case 2 scheme could be considered as a good process option for the economic purpose.

CHAPTER 4. Concluding Remarks

4.1. Conclusions

Unlike conventional feed compositions, this thesis focused on lean gas feed conditions, and proposed simplified NGL recovery as well as NGL/LNG integrated processes. These proposed processes were compared process efficiency and economy performance with the traditional representative processes. The scrub column process shows the best overall economic performance, considering both the total capital cost and operating cost, compared with other representative NGL recovery processes, such as ISS and IPSI, when the feed GPM value is lower than 2.3. For the NGL/LNG integrated processes, the proposed integrated process case 1 scheme presents lower total capital cost than the base case process. Case 1 scheme also shows lower total operating cost and TAC compared with the base case and case 2 processes under the lean feed condition. Although the proposed case 1 process shows advantages in total capital cost and operating cost relative to the base case, it maintained similar performance compared with the base case process in terms of the liquefaction efficiency and LNG production.

The results demonstrate that the proposed processes have advantages on overall economic performance compared with the conventional processes with relatively low loss of process efficiency under the lean feed conditions. Therefore, the proposed processes could be seen as another consideration for selecting a NGL recovery and NGL/LNG integrated processes when the feed composition is lean.

Moreover, various types of liquefaction processes for offshore units were investigated considering both liquefaction efficiency and economic performance.

The results provide a quantitative analysis of both the liquefaction efficiency and economics of those liquefaction processes that will be an important reference when selecting a liquefaction process for offshore application.

4.2. Future works

We find that the method of calculating the major equipment cost used in this study tends to underestimate in some special items, such as compressors and turbines, compared to the actual price when collaborative research with engineering companies. Therefore, further studies need to improve the method of calculating equipment purchase cost.

In the case of the operating costs, there is a limitation that are overestimated in terms of net profit, since only utility costs and raw materials were considered as the operating costs in this study. Further research should be needed to assess other factors of operating costs such as fixed charges and administrative costs for the NGL/LNG integrated processes.

Economic evaluation and process optimization were performed for the proposed integrated processes in this dissertation. However, the industrial operability for the proposed processes should be further evaluated.

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Nomenclature

Abbreviations

AGRU	Acid gas removal unit
APCI	Air products and chemicals
API	American petroleum institute
C3MR	Propane pre-cooled mixed refrigerant
CEPCI	Chemical engineering plant and cost index
CMR	Cold mixed refrigerant
CRR	Cold residue recycle
DMR	Dual mixed refrigerant
EIA	U.S Energy Information Administration
ESDU	Engineering Sciences Data Unit
GA	Genetic algorithm
GPM	Gallons of liquid per thousand standard cubic feet
GPSA	Gas processors suppliers association
GSP	Gas sub-cooled
HETP	Height equivalent to theoretical plate
HHC	Heavy hydrocarbons
HHV	Higher heating value
IGU	International gas union
ISS	Industry standard single stage
JT	Joule-Thomson
LNG	Liquefied natural gas
LNG FPSO	Floating production storage and off-loading units
MFC	Mixed fluid cascade
MR	Mixed refrigerant
MTPA	Million tons per annum
NG	Natural gas
NGL	Natural gas liquids
PMR	Precooling mixed refrigerant
SMR	Single mixed refrigerant

TAC	Total annualized cost
TCC	Total capital cost
WMR	Warm mixed refrigerant

Abstract in Korean (국문초록)

최근 호주와 미국을 비롯한 국가들의 비 전통가스전 개발이 증가함에 따라 lean 가스필드에 대한 수요가 점차 증가하고 있으며 따라서 조성이 lean 한 조건하에서의 NGL 회수공정 및 LNG 공정에 대한 성능 및 경제성 재평가가 필요할 것으로 예상된다. 본 논문은 다양한 NGL 회수 공정과 액화공정 그리고 NGL/LNG 통합 공정에 대하여 HHV 스펙을 고려한 공정최적화 및 경제성평가를 진행하고 비교 분석하였다.

NGL회수공정은 네 가지 서로 다른 공정들에 대하여 다양한 lean가스조성 하에서 공정성능 및 경제성평가를 진행하였다. 그 중 ISS 와 IPSI 공정은 전통 feed 조건하에서의 대표적인 공정인 반면에 HHC separator 와 scrub column 은 장치수를 최소화하여 단순한 공정도를 가진 공정들로 feed 조성이 lean 할 경우 강점을 가질 것으로 예상되는 공정이다. 공정성능평가 결과 비교적 많은 장치를 사용하여 복잡한 공정도를 가지고 있는 IPSI 공정이 가장 좋은 분리효율을 가지고 있으므로 다른 공정들에 비해 가장 적은 재료비를 사용하였다. 그러나 상대적으로 높은 자본투자가 IPSI 공정의 전체적인 경제성에 더 큰 영향을 주었다. 가장 단순한 공정인HHC separator 공정은 다른 공정들 대비 가장 적은 자본투자비를 보였지만 상대적으로 좋지 않은 분리효율로 인하여 가장 많은 재료비를 사용하였다. 자본투자비용 및 운전비용 등 전반적인 경제성을 고려

했을 경우 ISS 공정이 feed GPM 값이 2.5로 근접할 때 가장 좋은 경제성을 보였고 scrub column 공정은 feed 조성이 일정하게 lean 할 경우 다른 공정들 대비 가장 좋은 경제성을 보였다. 이는 전반적인 경제성측면에서 봤을 때 조성이 일정하게 lean 할 경우에는 상대적으로 복잡한 ISS 공정이나 IPSI 공정보다도 scrub column 공정을 NGL 회수 공정으로 사용하는 것이 더 유리함을 보여 준다.

해상용 천연가스 액화공정의 경우 제한된 공간 및 안전성 등 원인으로 인하여 일반적으로 육상보다 더 복잡한 선정기준을 가지고 있다. 예를 들어 육상에서 최대 점유율을 보이고 있는 C3MR 액화공정은 상대적으로 많은 공간 필요 및 공정안전성에 영향을 주는 많은 양의 프로판 성분을 필요로 하고 있으며 이로 인하여 해양 플랫폼에는 실제 사용 된 경우가 없다. 본 논문은 해상에서 사용 가능한 하나의 N2 엑스펜다 공정, 두 가지 종류의 단일혼합냉매 공정 (SMR) 및 세가지 타입의 듀얼혼합냉매 공정 (DMR) 등 총 6가지 타입의 공정들에 대해 공정성능 및 경제성평가를 진행하였다. N2 엑스펜다 공정은 비가연성인 질소를 단일냉매로 사용하므로 플랜트 운전측면 및 안전성 측면에서 강점을 가지고 있지만 다른 공정들 대비 가장 좋지 않은 효율을 보이는 것으로 알려져 있다. DMR 공정은 두 개의 혼합냉매를 사용하므로 N2 엑스펜다 및 SMR 공정보다 더 좋은 효율을 보이고 트레인당 용량도 가장 크므로 상대적으로 큰 용량의 액화공정 선정에 사용가능하다. 반면에 상대적으로 많은 장치를 사용하므로 복잡한 공정도를 가지고 있고 따라서 가장 큰 투자비를

필요로 한다. SMR 공정은 N2 엑스펜도 와 DMR 공정의 사이의 공정성능을 보인다. 성능 및 경제성평가 결과 N2 엑스펜더 공정이 다른 공정들 대비 가장 낮은 효율과 수익성을 보였다. SMR 2 공정은 가장 적은 투자비와 payout time 을 보였고 DMR base 공정이 다른 공정들 대비 높은 액화효율을 보여주었고 따라서 가장 높은 수익성과 가장 적은 운전비용을 필요로 하였다. SMR 2 공정은 투자비측면에서 가장 좋은 경제성을 보였고 DMR base 공정은 상대적으로 복잡한 공정도를 가지고 있어 높은 초기 투자비를 보이지만 높은 액화효율을 가지고 있어 운전비용 측면에서 다른 공정들 대비 강점을 가지고 있으므로 해상용 액화공정 선정 시 프로젝트 상황에 따라 SMR 2 공정 또는 DMR base 공정을 액화공정 후보로 고려할 수 있다.

통합공정은 본 논문에서 제안 한 두 개의 최대한 장치수를 간소화 한 NGL/LNG 통합공정과 전통적인 통합공정을 feed 가 lean 한 조성 하에서 LNG HHV 스펙을 고려하여 성능 및 경제성을 비교 분석하였다. 액화공정 파트는 SMR 2 공정과 DMR base 공정을 각각 사용하였고 유전자 알고리즘을 공정 글로벌 최적화에 적용하여 공정 최적운전조건을 도출하였다. 공정 최적화 결과 제안 한 Case 1 공정이 전통공정에 대비 액화효율은 조금 낮았지만 NGL회수 공정부분 장치를 최대한 줄임으로 인하여 상당히 낮은 자본투자비를 보였다. 또한 전반적인 경제성 측면에서도 Case 1 공정이 가장 좋은 성능을 보여주었다. 액화공정에 SMR 를 사용한 Case 2 공정의 경우에는 비교 분석한 세 개의 공정 중에서 가장 낮은 자본투자비를 보였으나 운전비

측면에서 다른 공정들 대비 비교적 큰 차이를 보여 전반적인 경제성 측면에서는 좋지 않았다. 하지만 플랜트 운영기간이 15개월보다 짧을 경우에는 가장 좋은 수익성을 보여주었다. 그러므로 제안된 Case 2 공정은 플랜트 운전기간이 짧은 peak shaving 플랜트 또는 feed reservoir 수명이 짧은 해양플랜트에 적용하면 전체 투자비를 줄일 수 있을 뿐만 아니라 전체 수익성 측면에서 유리하여 NGL/LNG 통합 공정 선정 시 하나의 좋은 공정옵션으로 고려할 수 있음을 보여주었다.

주요어: 경제성평가, NGL 회수, 액화공정, lean 가스

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